PB85-155695

Cost Digest: Cost Summaries of Selected Environmental Control Technologies

Radian Corp., Austin, TX

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

Oct 84

U.S. Department of Commerce
National Technical Information Service
NTIS

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/8-84-010	2.	3. RECIPIENT'S ACCESSION NO. PB85 155695/AS
4. TITLE AND SUBTITLE The Cost Digest--Cost Summaries of Selected Environmental Control Technologies		5. REPORT DATE Oct. '84
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Glen DeWolf, Pat Murin, James Jarvis, and Mary Kelly		8. PERFORMING ORGANIZATION REPORT NO. DCN 82-203-001-47-03
9. PERFORMING ORGANIZATION NAME AND ADDRESS Radian Corporation P.O. Box 9948 Austin, Texas 78766		10. PROGRAM ELEMENT NO.
		11. CONTRACT/GRANT NO. 68-02-3171, Task 47
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED User Manual; 1/81 - 6/82
		14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is John O. Milliken, Mail Drop 62, 919/541-7716.

16. ABSTRACT The report summarizes cost data on over 20 environmental control technologies. The cost parameters presented include total capital investment, net annual operating expenses, and unit annualized costs. These cost estimates are given over an appropriate range of system capacities for six technologies in drinking water treatment (e.g., granular activated carbon), eight technologies in wastewater treatment (e.g., conventional secondary and advanced), four particulate matter control technologies (e.g., fabric filters), and six flue gas desulfurization technologies (e.g., lime/limestone scrubbing). For each technology, the report contains a short narrative summary, a block flow diagram with battery limits noted around the modules included in the cost estimate, design specifications, performance characteristics, and a discussion of major factors affecting cost. All costs are given in March 1980 dollars.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Aerosols	Pollution Control	13B	07D
Cost Effectiveness	Flue Gases	Stationary Sources	14A	21B
Water Treatment	Desulfurization	Particulate		07A
Potable Water				
Waste Water				
Dust			11G	

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 105
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

PB85-155695

United States
Environmental Protection
Agency

Office of Environmental Engineering
and Technology
Washington DC 20460

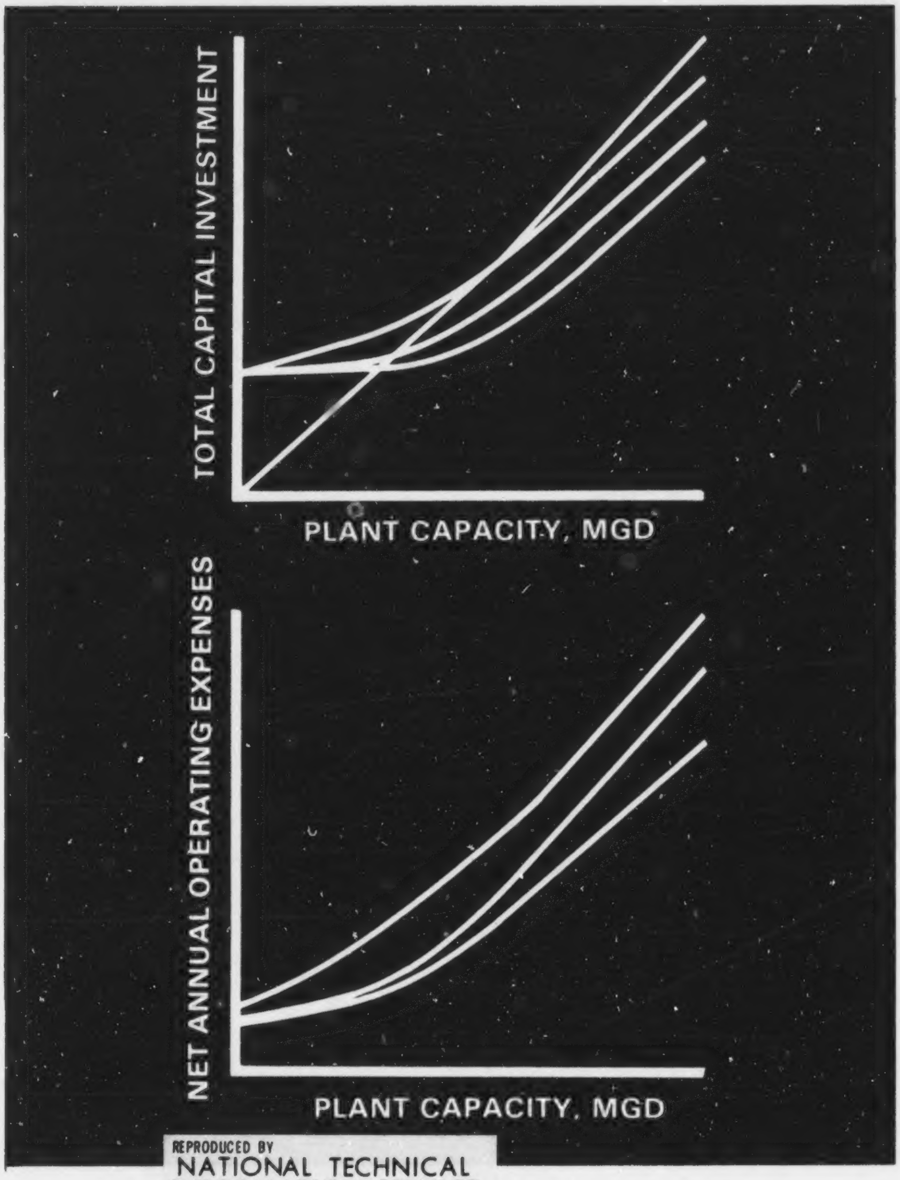
EPA-600/8-84-010
October 1984

Research and Development

EPA

The Cost Digest:

Cost Summaries of Selected Environmental Control Technologies



REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

EPA-600/8-84-010
October 1984

The Cost Digest:

Cost Summaries of Selected Environmental Control Technologies

by

Glenn DeWolf, Pat Murin, James Jarvis, and Mary Kelly

Radian Corporation
Austin, TX 78766

EPA Contract No. 68-02-3171

EPA Project Officer
John Milliken
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Office of Environmental Engineering and Technology
Office of Research and Development
U.S. Environmental Protection Agency
Washington, DC 20460

1a

Notice

This document has been reviewed in accordance with U.S. Environmental Protection Agency policy and approved for publication. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

Acknowledgements

This work was supported by the Office of Environmental Engineering and Technology Work Group on Costs of Environmental Control Technologies under the overall guidance of Steven Reznek (formerly ORD, Washington) and Kurt Riegel (ORD, Washington). Gene Tucker (IERL, RTP) was chairman of the work group supported by members John Milliken (IERL, RTP) (Project Officer), Alden Christianson (IERL, Cincinnati), Robert Clark (MERL, Cincinnati), Albert Klee (IERL, Cincinnati), Robert Smith (MERL, Cincinnati, now retired), and Richard Eilers (MERL, Cincinnati). In addition, other EPA staff members provided helpful suggestions and comments.

This report was prepared by Glenn DeWolf, James Jarvis, Pat Murin, and Mary Kelly of Radian Corporation with support and contributions from other staff members. It was prepared in fulfillment of Contract No. 68-02-3171, Work Assignments 26 and 47.

Preface

Environmental policy planners, permit writers and reviewers, management and budget officials, and developers of environmental control technologies use cost information on environmental control technologies to make effective decisions. Reliable, concise, and understandable cost data on capital investment, operating expenses, and revenue requirements serve to reduce the manager's level of uncertainty, and consequently improve his overall performance in attainment of environmental policy goals.

To provide reliable, concise, and understandable cost data, EPA's Office of Environmental Engineering and Technology (OEET) presents *THE COST DIGEST* as the first report in a series of publications on costs of environmental control technologies. This volume provides summary cost data for 25 selected environmental control technologies in the following areas: the treatment of drinking water and wastewaters, and the control of airborne particulate matter and sulfur oxides from stationary sources. In addition to cost data on capital investment and operating expenses for each technology, we have given special attention to providing facility design descriptions and control technology performance characteristics. These technology descriptions feature a narrative summary, a process flow chart with battery limits which illustrate the modules included in the cost estimates, key design parameters, and performance characteristics. The major variables affecting costs for each technology are also discussed. Although we have attempted to select representative or typical design configurations for each technology, the information on design parameters and performance characteristics is essential to effective use of associated cost data.

Two additional publications for the OEET series on costs of environmental control technologies are currently in preparation and review. These are *"COSTS OF ENVIRONMENTAL CONTROL TECHNOLOGIES—GRANULAR ACTIVATED CARBON APPLICATIONS IN WATER AND WASTEWATER TREATMENT"* and *"COSTS OF ENVIRONMENTAL CONTROL TECHNOLOGIES—PARTICULATE MATTER CONTROL FOR INDUSTRIAL AND UTILITY BOILERS."* Instead of presenting summary cost information for complete control technology systems, these volumes will provide more detailed engineering cost data for the specific modules which make up the control technology systems. This feature will allow for cost estimates to be more tailored to specific cases. By contrast, *THE COST DIGEST* allows the user to derive costs of typical, but mostly fixed, designs for control systems.

It is hoped that, as an executive summary of environmental technology cost information, *THE COST DIGEST* will be widely used by planners, budgeters, technology developers, and managers in general who need quick reference to easy-to-use, reliable cost data. We welcome comments on *THE COST DIGEST* and suggestions for guiding and improving the OEET reports on costs of environmental technologies.

Contents

Page

Acknowledgements iii

Preface iv

1.0 Introduction and User Guide 1
  1.1 Organization of the Report 1
  1.2 Terminology and Format for Presenting Cost Estimates 2
  1.3 Cost Updating 4
  1.4 Sources of Information and Limitations 4
  1.5 Considerations When Comparing Cost Estimates 5
  1.6 Relating Costs to Consumer Prices 6
  References 7

2.0 Drinking Water Treatment 8
  2.1 Filtration Treatment Plants 10
    2.1.1 Description 10
    2.1.2 Design Basis and Costs 11
    2.1.3 Major Variables Affecting Costs 12
  2.2 Disinfection 13
    2.2.1 Chlorine 13
    2.2.2 Chlorine Dioxide 14
    2.2.3 Ozone 16
    2.2.4 Chloramination 18
  2.3 Granular Activated Carbon Treatment 19
  2.4 Aeration 22
  References 24

3.0 Wastewater Treatment 26
  3.1 Conventional Secondary and Advanced Wastewater Treatment 27
  3.2 Stabilization Ponds and Aerated Lagoons 33
  3.3 Land Treatment 36
  3.4 Phosphorus Removal by Chemical Addition 39
  3.5 Nitrification (Separate-Stage) 40
  3.6 Granular Media Filtration 43
  References 45

4.0 Particulate Matter Collection 47
  4.1 Multitube Cyclones 49
  4.2 Electrostatic Precipitators 50
  4.3 Fabric Filters 55
  4.4 Venturi Wet Scrubbers 58
  References 62

Contents (continued)

Page

5.0 Flue Gas Desulfurization 64
  5.1 Wet FGD Scrubbing Processes 64
    5.1.1 Wet FGD Process Descriptions 64
    5.1.2 Design Basis and Costs 68
    5.1.3 Major Variables Affecting Costs 71
    5.1.4 Utility Boiler FGD Systems 71
    5.1.5 Industrial Boiler FGD Systems.................... 74
  5.2 Lime Spray Drying Process 75
    5.2.1 Process Description.............................. 75
    5.2.2 Design Basis and Costs 77
    5.2.3 Major Variables Affecting Costs 77
    References 80

Appendix A — Methods for Adjusting Data 82
  A.1 Format, Cost Factors, and Unit Prices 82
  A.2 Unit Annualized Cost Calculations 83
  A.3 Updating Costs........................... 86
  A.4 Interest During Construction 86
  A.5 Location Factors......................... 87
  References 87

Appendix B — Glossary 89
Appendix C — Conversion of English to International System (SI) Units 91
Appendix D — Miscellaneous Conversion Factors 92

List of Figures

Figure Page

Section 2

2-1 Conventional filtration system for drinking water treatment 10
2-2 Filtration plants for drinking water treatment
Total capital investment (March, 1980 dollars) 12
2-3 Filtration plants for drinking water treatment
Net annual operating expenses (March, 1980 dollars)........... 12
2-4 Filtration plants for drinking water treatment
Unit annualized cost (March, 1980 dollars) 12
2-5 Chlorination system for drinking water treatment 13
2-6 Chlorination system for drinking water treatment
Total capital investment (March, 1980 dollars) 14
2-7 Chlorination system for drinking water treatment
Net annual operating expenses (March, 1980 dollars)........... 14
2-8 Chlorination system for drinking water treatment
Unit annualized cost (March, 1980 dollars) 14
2-9 Chlorine dioxide system for drinking water treatment 15
2-10 Chlorine dioxide system for drinking water treatment
Total capital investment (March, 1980 dollars) 15
2-11 Chlorine dioxide system for drinking water treatment
Net annual operating expenses (March, 1980 dollars)........... 16
2-12 Chlorine dioxide system for drinking water treatment
Unit annualized cost (March, 1980 dollars) 16
2-13 Ozonation system for drinking water treatment 17
2-14 Ozonation system for drinking water treatment
Total capital investment (March, 1980 dollars) 17
2-15 Ozonation system for drinking water treatment
Net annual operating expenses (March, 1980 dollars)........... 17
2-16 Ozonation system for drinking water treatment
Unit annualized cost (March, 1980 dollars) 18
2-17 Chloramination system for drinking water treatment 18
2-18 Ammonia feed system for drinking water treatment by chloramination
Total capital investment (March, 1980 dollars) 19
2-19 Ammonia feed system for drinking water treatment by chloramination
Net annual operating expenses (March, 1980 dollars)........... 19
2-20 Ammonia feed system for drinking water treatment by chloramination
Unit annualized cost (March, 1980 dollars) 20
2-21 Granular activated carbon system for drinking water treatment ... 20

List of Figures

Figure Page

2-22 Granular activated carbon system for drinking water treatment
Total capital investment (March, 1980 dollars) 21

2-23 Granular activated carbon system for drinking water treatment
Net annual operating expenses (March, 1980 dollars).......... 21

2-24 Granular activated carbon system for drinking water treatment
Unit annualized cost (March, 1980 dollars) 22

2-25 Aeration systems for drinking water treatment 23

2-26 Aeration for drinking water treatment
Total capital investment (March, 1980 dollars) 24

2-27 Aeration for drinking water treatment
Net annual operating expenses (March, 1980 dollars).......... 25

2-28 Aeration for drinking water treatment
Unit annualized cost (March, 1980 dollars) 25

Section 3

3-1 Conventional secondary treatment system for wastewater 28

3-2 Advanced wastewater treatment system 29

3-3 Conventional secondary and advanced wastewater treatment
Total capital investment (March, 1980 dollars) 32

3-4 Conventional secondary and advanced wastewater treatment
Net annual operating expenses (March, 1980 dollars).......... 32

3-5 Conventional secondary and advanced wastewater treatment
Unit annualized cost (March, 1980 dollars) 33

3-6 Stabilization pond or aerated lagoon system for wastewater
treatment 34

3-7 Stabilization ponds and aerated lagoons for wastewater
treatment
Total capital investment (March, 1980 dollars) 35

3-8 Stabilization ponds and aerated lagoons for wastewater
treatment
Net annual operating expenses (March, 1980 dollars).......... 35

3-9 Stabilization ponds and aerated lagoons for wastewater
treatment
Unit annualized cost (March, 1980 dollars) 36

3-10 Land treatment system for wastewater 37

3-11 Land treatment for wastewater treatment
Total capital investment (March, 1980 dollars) 38

3-12 Land treatment for wastewater treatment
Net annual operating expense (March, 1980 dollars) 38

3-13 Land treatment for wastewater treatment
Unit annualized cost (March, 1980 dollars) 38

3-14 Phosphorus removal by chemical addition for wastewater
treatment 39

3-15 Phosphorus removal for wastewater treatment
Total capital investment (March, 1980 dollars) 40

3-16 Phosphorus removal for wastewater treatment
Net annual operating expenses (March, 1980 dollars).......... 41

List of Figures

Figure Page

3-17 Phosphorus removal for wastewater treatment Unit annualized cost (March, 1980 dollars) 41

3-18 Separate-stage nitrification system for wastewater treatment 42

3-19 Nitrification (separate-stage) for wastewater treatment Total capital investment (March, 1980 dollars) 43

3-20 Nitrification (separate-stage) for wastewater treatment Net annual operating expenses (March, 1980 dollars)........... 43

3-21 Nitrification (separate-stage) for wastewater treatment Unit annualized cost (March, 1980 dollars) 43

3-22 Granular media filter system for wastewater treatment 44

3-23 Granular media filtration for wastewater treatment Total capital investment (March, 1980 dollars) 45

3-24 Granular media filtration for wastewater treatment Net annual operating expenses (March, 1980 dollars)........... 45

3-25 Granular media filtration for wastewater treatment Unit annualized cost (March, 1980 dollars) 45

Section 4

4-1 Illustration of collection efficiency versus particle diameter 48

4-2 Multitube cyclone system for particulate matter collection 49

4-3 Multitube cyclone system for particulate matter collection Total capital investment (March, 1980 dollars) 50

4-4 Multitube cyclone system for particulate matter collection Net annual operating expenses (March, 1980 dollars)........... 51

4-5 Multitube cyclone system for particulate matter collection Unit annualized cost (March, 1980 dollars) 51

4-6 Electrostatic precipitator system for particulate matter collection .. 52

4-7 Electrostatic precipitator system for particulate matter collection Total capital investment (March, 1980 dollars) 54

4-8 Electrostatic precipitator system for particulate matter collection Net annual operating expenses (March, 1980 dollars)........... 54

4-9 Electrostatic precipitator system for particulate matter collection Unit annualized cost (March, 1980 dollars) 55

4-10 Fabric filter system for particulate matter collection 56

4-11 Fabric filter system for particulate matter collection Total capital investment (March, 1980 dollars) 57

4-12 Fabric filter system for particulate matter collection Net annual operating expenses (March, 1980 dollars)........... 57

4-13 Fabric filter system for particulate matter collection Unit annualized cost (March, 1980 dollars) 58

4-14 Venturi wet scrubber system for particulate matter collection..... 59

4-15 Venturi wet scrubber comparative fractional efficiency curves 60

4-16 Venturi wet scrubber system for particulate collection Total capital investment (March, 1980 dollars) 60

4-17 Venturi wet scrubber system for particulate collection Net annual operating expenses (March, 1980 dollars)........... 61

List of Figures

Figure **Page**

4-18 Venturi wet scrubber system for particulate collection Unit annualized cost (March, 1980 dollars) 61

Section 5

5-1 Lime/limestone scrubbing process for flue gas desulfurization 65

5-2 Sodium alkali scrubbing (throwaway) process for flue gas desulfurization 66

5-3 Dual alkali scrubbing process for flue gas desulfurization 67

5-4 Magnesium oxide scrubbing process for flue gas desulfurization .. 68

5-5 Wellman-Lord scrubbing process for flue gas desulfurization 69

5-6 Flue gas desulfurization systems for utility boilers Total capital investment (March, 1980 dollars) 71

5-7 Flue gas desulfurization systems for utility boilers Net annual operating expenses (March, 1980 dollars)........... 71

5-8 Flue gas desulfurization systems for utility boilers Unit annualized cost (March, 1980 dollars) 72

5-9 Flue gas desulfurization systems for industrial boilers Total capital investment (March, 1980 dollars) 72

5-10 Flue gas desulfurization systems for industrial boilers Net annual operating expenses (March, 1980 dollars)........... 73

5-11 Flue gas desulfurization systems for industrial boilers Unit annualized cost (March, 1980 dollars) 73

5-12 Lime spray drying process for flue gas desulfurization 76

5-13 Lime spray drying flue gas desulfurization systems for utility boilers Total capital investment (March, 1980 dollars) 78

5-14 Lime spray drying flue gas desulfurization systems for utility boilers Net annual operating expenses (March, 1980 dollars)........... 78

5-15 Lime spray drying flue gas desulfurization systems for utility boilers Unit annualized cost (March, 1980 dollars) 79

5-16 Lime spray drying flue gas desulfurization systems for industrial boilers Total capital investment (March, 1980 dollars) 79

5-17 Lime spray drying flue gas desulfurization systems for industrial boilers Net annual operating expenses (March, 1980 dollars)........... 80

5-18 Lime spray drying flue gas desulfurization systems for industrial boilers Unit annualized cost (March, 1980 dollars) 80

List of Tables

Table Page

Section 1

1-1 Technology Areas Addressed 1
1-2 Format for Total Capital Investment 2
1-3 Format for Net Annual Operating Expenses 2

Section 2

2-1 Drinking Water Contaminant Levels Based on Safe Drinking Water Act 9
2-2 Charge Rate Profile for Typical Total Drinking Water System Costs Based on Survey Data 9
2-3 Design Parameters for Typical Granular Activated Carbon Systems for Drinking Water Treatment 21
2-4 Major Design Parameters for Aeration Basins and Towers for Trihalomethane Removal in Drinking Water Treatment 23

Section 3

3-1 Typical Pollutant Removal Efficiency of Wastewater Treatment Technology 26
3-2 Typical Influent Wastewater Composition 30
3-3 Design Parameters for Conventional Secondary Treatment System 31
3-4 Design Parameters for Advanced Wastewater Treatment System 31
3-5 Design Parameters for Pond Systems 34
3-6 Design Parameters for Land Treatment 37
3-7 Design Parameters for Nitrification (Separate-Stage) 42
3-8 Design Parameters for Granular Media Filtration 44

Section 4

4-1 Comparison of Major Particulate Collection Systems 47
4-2 Comparison of Wet and Dry Collection Systems 48
4-3 Required Collection Efficiency for Typical Uncontrolled and Hypothetical Controlled Particulate Matter Concentrations 48
4-4 Multitube Cyclone Design Parameters 50
4-5 Precipitation Rate Parameters for Typical ESP Applications 53
4-6 Design Parameters for Model Electrostatic Precipitators 53
4-7 Design Parameters for Model Fabric Filter Systems 57
4-8 Venturi Wet Scrubber Design Parameters 60

List of Tables

Table Page

Section 5

5-1 Comparison of Design Bases for Major Cost References 69
5-2 Effect of Changes in the Design Basis on PEDCo Total Capital Investment Estimates 74
5-3 Effect of Changes in the Design Basis on TVA Total Capital Investment Estimates 74
5-4 Effect of Coal Sulfur Content on Total Capital Investment and Total Annual Operating Expenses for Utility Boiler Applications ... 75
5-5 Effect of Coal Sulfur Content and SO_2 Removal Efficiency on Total Capital Investment and Total Annual Operating Expenses for Industrial Boiler Applications 76
5-6 Design Bases for Utility and Industrial Lime Spray Dryer/Fabric Filter Systems 77

Appendix A

A-1 Format and Factors for Total Capital Investment 83
A-2 Format for Net Annual Operating Expenses 84
A-3 Unit Prices Employed for Net Annual Operating Expenses 84
A-4 Basis for Fixed Charge Rate Annualized Cost Calculations 85
A-5 Annual Average and End-of-Quarter Capital Cost Indices 86
A-6 Factors for Calculating Interest During Construction 87
A-7 Cost Locality Factors 87
A-8 Power Cost Locality Factor 87

Section 1
Introduction and User Guide

Planners and managers in government and industry require cost information to aid in policy planning, implementation, and administration. Much cost information is available, but it is scattered in numbers of published sources and not readily accessible for quick reference. Furthermore, information from different sources varies in the cost bases used, format of presentation, level of detail, accuracy, documentation, and applicability for broad-based strategic planning and analysis.

This manual was prepared to provide a concise and easily understood graphical compilation of costs for selected environmental control technologies in the following areas:

- Drinking water treatment.
- Wastewater treatment (municipal and industrial).
- Particulate matter control.
- Flue gas desulfurization.

A further goal was to present cost data in a consistent format and terminology to allow ready interpretation without extensive analysis and calculations by the user. Finally, this publication was intended as a summary document which could be revised, updated, and augmented in order to keep pace with new developments in key environmental control technologies.

Each of the four study areas addressed in this report comprises several technologies as shown in Table 1-1. This list was selected by the Work Group on Environmental Control Technologies, Office of Environmental Engineering and Technology, within the Office of Research and Development. The technologies were chosen to represent those environmental control options in the four study areas currently of interest to policy planners.

Table 1-1. Technology Areas Addressed

- Drinking Water Treatment Systems (Section 2)
    - Filtration treatment (conventional filtration, direct filtration, and lime softening with conventional filtration)
    - Disinfection
    - Granular activated carbon treatment
    - Aeration
- Wastewater Treatment Systems (Municipal and Industrial) (Section 3)
    - Conventional secondary (less than 30 mg/l BOD$_5$) and advanced wastewater treatment plants (less than 10 mg/l BOD$_5$)[a]
    - Stabilization ponds and aerated lagoons
    - Land treatment
    - Phosphorus removal by chemical addition
    - Nitrification
    - Granular media filtration
- Particulate Matter Control Systems (Section 4)
    - Mechanical collectors (multitube cyclones)
    - Electrostatic precipitators
    - Fabric filters
    - Venturi scrubbers
- Flue Gas Desulfurization Systems (Section 5)
    - Lime/limestone scrubbing
    - Non-regenerable sodium alkali (throwaway)
    - Dual alkali
    - Magnesium oxide
    - Wellman-Lord
    - Dry scrubbing

[a] Conventional secondary treatment is defined as achievement of 30 mg/l biochemical oxygen demand (BOD$_5$) and 30 mg/l suspended solids (SS) or less in the treatment system effluent. Advanced wastewater treatment achieves 10 mg/l BOD$_5$ and 10 mg/l SS or less.

1.1 Organization of the Report

Sections 1.2 through 1.6 provide guidance in interpreting and using the cost data in this report. Section 1.2 describes the terminology and format used for presenting the cost data as well as the overall methodology for cost development. Section 1.3 discusses cost updating. Section 1.4 discusses the sources of data used in developing this manual and limitations to its use. Section 1.5, a brief discussion on general considerations when comparing cost estimates, is presented to give the reader some perspective in using the cost estimates in this report and/or comparing them with cost estimates in other references. Additional variables specific to individual technologies are discussed in the individual technology sections. Section 1.6 provides some examples that show how environmental control costs can be related to consumer prices.

The technology areas are discussed in Sections 2 through 5. Each section is divided into subsections for each individual technology. Each subsection presents a brief description highlighting the major technical features of the process and the design basis for costs presented. Graphical displays of total capital investment, net annual operating expenses, and unit annualized costs are provided. A discussion of major technology-specific variables affecting costs completes each technology subsection.

The appendices contain details which supplement material discussed in the main body of the report. Appendix A describes the methodology used to develop the costs presented for each technology area. Appendix B is a glossary of cost-related and technical terminology. The reader should confirm his/her familiarity with the cost terms to verify that definitions are consistent with the intended use of the information. Appendix C is a list of conversion factors for English and SI units, and Appendix D, miscellaneous conversion factors between units of measure.

1.2 Terminology and Format for Presenting Cost Estimates

Graphs are provided for each technology system showing total capital investment, net annual operating expenses, and unit annualized cost. These cost terms have a specific meaning and usage within the format discussed in a report by Uhl (1). A feature of that format is the assignment of an item number to the individual cost elements comprising each of the above three cost items. This numbering procedure is used to ensure unambiguous interpretation of cost elements in using the methodology of Uhl's report even if different authors use a different terminology to describe various cost items. Listings of the item numbers, cost elements for total capital investment, and cost elements for net annual operating expenses are given in Table 1-2 and Table 1-3, respectively. Unit annualized cost is computed from fixed capital charges and net annual operating expenses as explained below. For simplification in the present report, some individual cost elements were combined into a single overall category. A line item with several numbers next to it indicates that several individual elements have been combined.

Total Capital Investment

All capital costs in the present report are shown as total capital investment. Total capital investment comprises 40 numbered cost elements as shown in Table 1-2 and is itself designated as item 41. Various subtotals are shown in upper case letters in Table 1-2. Each succeeding subtotal is obtained by adding cost elements to the preceding subtotal. With the exception of direct cost items and land, all cost elements are determined by multiplying a subtotal by a factor. For example, a contingency allowance is obtained by multiplying total bare module cost by a factor. Factors used for each technology area are shown in Table A-1, Appendix A. The direct costs for each technology were adapted from costs in the technical literature as discussed in Appendix A.

Direct cost items include both installed purchased equipment and field fabricated process equipment. Pumps are an example of installed purchased equipment. Field fabricated process equipment includes such items as the concrete basins used in certain drinking water and wastewater treatment processes. Some references refer to direct costs as construction costs or installed equipment costs.

Table 1-2. Format for Total Capital Investment

Item No.[a]	Item	Cost[b]
1-10	Direct cost items	-[c]
11	TOTAL DIRECT COST	
12-20	Indirect cost items (Engineering and construction and field expenses, other)	(F_1 x Item 11)
21	TOTAL BARE MODULE COST	(Item 11 + Items 12 through 20)
22	Contingency	(F_2 x Item 21)
23	Contractor's fee	(F_3 x Item 21)
27	Retrofit increment	(F_4 x Item 21)
24-26, 28-30	Other	(F_x x Item 21)
31	TOTAL PLANT COST	(Item 21 + Items 22 through 30)
32	Interest during construction	(F_5 x Item 31)
33	Start-up	(F_6 x Item 31)
34	Other	(F_y x Item 31)
35	TOTAL DEPRECIABLE INVESTMENT	(Item 31 + Items 32 through 34)
36	Land	(Direct calculation of cost)
37	Working capital	(F_7 x Item 35)[d]
38-40	Other	(F_z x Item 35)
41	TOTAL CAPITAL INVESTMENT	(Item 35 + Items 36 through 40)

[a] For a detailed discussion of individual line items and item numbers see Uhl (1).

[b] F_1, F_2, etc., refer to factors for cost element or line item.

[c] All installed equipment costs are added to arrive at total direct cost for the system.

[d] Other methods are also possible. See Uhl (1).

Table 1-3. Format for Net Annual Operating Expenses

Item No.[a]	Item
53	Raw materials
56-58, 61	Labor[b]
59, 60	Materials[c]
62	Steam
63	Power (Electricity)
64	Compressed air
65	Water
66	Fuel
67	Waste disposal
68, 69	Other
70	PROCESSING EXPENSES (sum of Items 53 through 69)
74	Overhead (F_1 x labor items)
76	Insurance and property taxes (F_2 x TDI[d])
80	NET OPERATING COSTS
87	General expense (F_3 x TDI[d])
88-89	Other
e	NET ANNUAL OPERATING EXPENSES

[a] For a detailed discussion of individual line items and item numbers see Uhl (1).

[b] Includes operating direct labor, direct supervision, maintenance labor, and labor burden.

[c] Includes maintenance materials and operating supplies.

[d] TDI = total depreciable investment; see Table 1-2.

[e] No item number was provided for this line item.

When indirect costs such as engineering and supervision and construction field expenses are added to the direct costs, the total bare module cost is obtained. Contingency and contractor's fees added to the total bare module cost yield total plant or system cost. Two additional cost elements which sometimes are capitalized include loan interest during construction and start-up costs incurred during initial operation of a new facility. If these costs are not capitalized but are treated directly as an annual expense, the total plant cost and total depreciable investment are identical. In this report, however, construction interest and start-up costs are capitalized and added to the total plant cost to obtain total depreciable investment. To obtain the total capital investment, the cost of land and working capital is added to the total depreciable investment.

Net Annual Operating Expense

Net annual operating expense refers to direct cash expenses of operation and maintenance as well as indirect items including overhead, insurance and property taxes, and general expenses. Inclusion of depreciation, a non-cash expense, would produce total annual operating expenses rather than net annual operating expenses. The elements of net annual operating expenses are shown in Table 1-3.

Part of the net annual operating expenses is processing expenses. These expenses are commonly referred to as O&M or operating and maintenance expenses. Because the term O&M is not universally defined as comprising the same cost elements, its use has been avoided in this report. Cost elements that make up processing expenses are determined directly from operating requirements and corresponding unit prices. Items added to processing expenses to generate the net annual operating expenses can be obtained by multiplying a factor times another cost element. In this report overhead was obtained as a factor times labor cost; insurance and property taxes and general expenses as a factor times total depreciable investment. As with capital cost elements discussed previously, each line item is assigned a number. And, again for simplification in this report, some line items have been combined so that in some cases several numbers appear in the item number columns.

Unit Annualized Cost

The unit annualized cost is the annualized cost divided by the annual capacity of the process to yield cost per unit of capacity such as cents per thousand gallons or cents per kilowatt-hour. The annualized cost is the sum of net annual operating expenses and additional cost elements. The additional cost elements added to net annual operating expenses cover depreciation, cost of financing, and an allowance for income taxes. Annualized cost is equivalent to the minimum annual revenue requirement for the project. The unit annualized cost is, therefore, equivalent to the minimum unit annual revenue requirement or unit price for the pollution control service performed.

A common method for including depreciation and costs of financing is to use a capital recovery factor where these cost elements are lumped into a single number. In this report, a form of capital recovery factor called the fixed charge rate is used. Typical financing assumptions were used to develop the unit annualized cost. This method and the assumptions, discussed in Appendix A, account for depreciation, cost of financing, income taxes, and the effect of an investment tax credit lumped into a single number.

Data Presentation

Certain key features of the cost data presented in this report include:

- *Data are presented graphically. Total capital investment, net annual operating expenses, and unit annualized cost are plotted against a system capacity variable.* In some cases, multiple curves are shown on a graph to illustrate cost variations caused by major variables specific to a technology. For drinking water and wastewater treatment, costs are given as a function of plant capacity in millions of gallons per day (mgd). For particulate control systems, the cost data are plotted against actual cubic feet per minute of gas stream flow, fuel firing rate for fired process or industrial boiler equipment, and megawatt generating capacity for utility boilers. Because FGD systems are used primarily on industrial and utility boilers, megawatt generating capacity and fuel firing rate in Btu/hr are the major variables against which costs are plotted. These choices were based on common usage in existing cost references for these technology areas. Conversion factors between gas flow rate, megawatt generating capacity, and fuel firing rate are provided in Appendix D.
- *Cost data are presented for entire treatment systems rather than individual system components.* This permits the user to obtain a typical pollution control system cost without extensive computational and design exercises. References used as sources for cost data from which costs in this report were adapted provide greater detail on component costs, but require selection of system parameters, addition of individual component costs to obtain total system costs, and other calculations. These have been done for the user for a typical or representative design and application for each technology.
- *Costs presented for each technology are for a typical or representative design and application.* Site-specific factors will result in actual system costs that might vary significantly from the values reported here. Some of the reasons for these

variations are discussed in Subsection 1.5 of this Introduction and User Guide, as well as in each individual technology section.

- *All costs apply for new environmental control technology systems as they would be installed in new facilities.* The capital cost data might be applied to retrofit situations in which new pollution control systems are installed at existing facilities. However, retrofitted pollution control systems incur a cost penalty that is not considered in the cost data presented here. Little documented information is available concerning cost penalties for retrofit installations. Some retrofit costs have been reported as a much as 70 percent higher than the capital investment for a comparable new installation (1).

1.3 Cost Updating

All costs in this report are expressed in March 1980 dollars. Costs reported in the literature were updated using cost indices and March 1980 unit prices for labor, materials, electricity, and fuel.

Costs expressed in base year dollars may be adjusted to dollars for another base year by applying cost indices as shown in the following equation:

$$\text{new base year cost} = \text{old base year cost} \times \frac{\text{new base year index}}{\text{old base year index}}$$

Capital costs from existing publications were updated using this method. In most cases, the level of detail available in cost references suggested that an overall index should be applied to the total direct capital costs rather than to individual items making up the total direct costs. Two indices were used in this report. For drinking water and wastewater treatment systems, the Engineering News Record (ENR) Construction Cost Index was used. The Chemical Engineering (CE) Plant Cost Index was used for particulate matter control and flue gas desulfurization systems. Values for these indices by year are given in Table A-5, Appendix A.

For March 1980 these indices are:

ENR Construction Cost Index	3150
CE Plant Cost Index	253

Most major cost components of net annual operating expense were updated individually using unit prices for March 1980. Tabulations of unit prices are given in Table A-3 in Appendix A. Costs for electricity and fuel were obtained from the Monthly Energy Review published by the Department of Energy (2). Materials' costs were updated using the Producer Price Index for Finished Goods. The Producer Price Index is used in the same way as the capital cost indices discussed above and was obtained from the Monthly Labor Review published by the Department of Labor (3). A basic labor rate was also obtained from this reference and adjusted upward for fringe benefits by applying a factor.

1.4 Sources of Information and Limitations

The costs presented in this report are derived from cost information in existing published sources. It was the objective of this report to prepare a cost summary for each technology using the best documented costs from the literature and to adjust these to a consistent basis. It was not the objective to generate new fundamental cost data. The primary sources of information are recent EPA publications supplemented by other references where necessary. System design, system boundaries (scope), format of data presentation, terminology, reference year, and unit cost values are variable between the different references. Adjustments were made to bring the data into a standard format as well as to update all costs to a March 1980 dollar basis. In addition, for some technologies, well documented system costs were not available so that they had to be developed from component costs.

A limitation of some of the cost literature is that explicit definitions of design bases are not always available. There is therefore an element of uncertainty in the scope and specifications for some of the cost data that have been used. Design bases in this report are stated as clearly and completely as the published information allowed. For each technology, design criteria are described and a table of key design parameters is presented where appropriate.

Costs in this document reflect the 'typical' or 'average' representation of specific technologies. This restricts the use of the data in this report to:

- Preliminary estimates used for policy planning.
- Comparison of relative costs of different technologies.
- Approximations of costs that might be incurred for a specific application.

The costs in this report are considered to be 'order of magnitude' with a ±50 percent margin. This is because cost curves are drawn based on updates and adjustments to literature costs for three or four system capacities for each technology. Large departures from the design basis of a technology in this report might cause the system costs to vary by a greater extent than this. If used as intended, however, this document will provide a reliable source of preliminary cost information for the technology areas covered.

When comparing costs in this report to costs from other references, the user should be sure the design bases are comparable and that total capital investment, net annual operating expenses, and unit annualized costs are actually the costs being compared. For example, O&M costs in many references are only part of the net annual operating expenses as used here.

1.5 Considerations When Comparing Cost Estimates

Two important considerations affecting cost estimates for any system are:

- design basis
- accounting methods (i.e., methodology).

These two factors probably have as much of an effect on apparent differences in estimated costs (and reported actual costs of completed projects) as any other factors.

Other factors which result in differences in reported costs are terminology and fundamental cost data such as item prices. Sources of published cost information do not always use the same terms to describe costs and do not always report costs at the same level of development. For example, in the list of Table 1-2, the term capital cost might be used to describe any of the items 11, 21, 31, 35, or 41 depending on individual interpretation. This problem occurs with operating costs as well as capital costs. Finally the differences in prices used for capital equipment and materials and unit prices for direct operating cost elements such as labor and power influence the results.

The design basis defines both the scope of a facility and specifications for the inidividual components comprising the facility. These determine the direct costs for the physical plant as well as indirect costs which typically are estimated as a percentage of direct costs. Cost elements of operating expenses such as labor and power requirements follow directly from the design basis since they are related to equipment design and operating requirements. Again, the indirect cost elements comprising operating expenses are dependent on the design basis because they are typically computed as a percentage of both capital costs and direct operating expenses. The prices used for various capital equipment and operating expense items, of course, influence the final result, but the quantities to which the prices are applied depend on system design.

A second major reason for differences in reported costs is costing methodology. This includes the selection of methods for calculating various subtotals of cost elements which, when added together, yield the desired cost total. Sometimes every cost element is estimated independently. Sometimes certain cost elements are derived from others. For estimating capital costs, a sequence of factors is commonly applied to purchased equipment costs or installed equipment costs to generate a total capital requirement. The terminology and level of summation at which the estimating procedure is terminated determine the cost values ultimately reported. Some insight into this aspect of estimate preparation is found in many literature sources (1). Similar considerations apply to net annual operating expense.

Reported experienced costs for actual completed facilities frequently differ from average estimated costs used for conceptual estimating. This difference is usually attributed to "site-specific factors." Sometimes the differences occur due to differences in cost accounting and the allocation of costs to specific categories. In other cases the site-specific factors are variables that legitimately influence costs and are highly specific to a particular facility.

Some of these site-specific factors are due to differences in individual waste source characteristics which give rise to differences in treatment system design. The design differences result in different costs for a system, even at the same level of performance, so that there is not always a simple direct relationship between performance and cost. The site-specific design which influences direct costs combined with many indirect cost considerations specific to a given project ultimately determines the cost for a particular facility.

Factors that may vary with individual projects noted by other authors as affecting costs include (4, 5):

- Competition in contractor and material supplier markets (i.e., business climate) resulting in unusually high or low bids and prices.
- Variations in local material and labor costs.
- Timing of construction with regard to the season of the year, length of construction period, and interest rates.
- Variations in conventional engineering, design, and construction practices.
- Special considerations superimposed on normal design requirements by local regulatory agencies.
- Cost consciousness and consideration given to cost control during design and construction.
- Physical and climatic variations in individual site conditions.
- Architectural features.

This discussion has highlighted some major cost-influencing factors common to all technologies. Additional discussion of some technology specific variables affecting costs is provided in the individual technology sections.

As discussed earlier, costs presented in each of the individual technology sections that follow are based on data from existing publications. Adjustments have been made so that the costs conform to the format and terminology discussed in this section and in Appendix A to this report. As explained above, each treatment technology addressed may have variations in the choice of equipment and the layout of the equipment comprising the system which will affect costs. In the existing cost literature for these technologies a complete definition of design scope and specifications is not always available. Within the constraints of existing literature, the costs presented here are an attempt to pro[illegible]he user of this report with a thorough representation of cost estimates that can currently be obtained for the selected technologies.

1.6 Relating Costs to Consumer Prices - Examples

One use of this report might be to provide information for a preliminary evaluation of cost impacts of environmental control technologies. A typical cost impact would be the effect on consumer prices.

Several examples are provided here that present costs of environmental control technologies in the perspective of the consumer. Treatment costs are related to a typical monthly consumer expenditure for a commodity which would require the treatment technology in its use or manufacture.

For drinking water and wastewater treatment in municipal applications, an example is given relating the cost to a typical monthly household billing for water service. A single example is used since the principles are the same in both of these technology areas. For flue gas desulfurization applied to a steam electric generating plant, the example shows the cost impact on the monthly electric bill. Finally, particulate matter control costs are related to the consumer price of a building product.

Municipal wastewater treatment costs can be related to typical household wastewater charges. Assume a household that discharges a total of 5000 gallons a month.* Using the unit annualized cost for any wastewater technology discussed in the subsections that follow, one can obtain a generalized average monthly cost of the treatment technology to the consumer. One multiplies the wastewater generated in 1000s of gallons by the appropriate unit annualized cost in dollars per 1000 gallons. Using the unit annualized cost of $1.00 per 1000 gallons for conventional secondary treatment plants from Figure 3-5 in this report (Section 3), the monthly charges to cover treatment would be $13.50 in a community of 70,000 people. If an advanced wastewater treatment plant were used, the unit annualized cost would be $1.80/1000 gallons. Using the typical household discussed above, the monthly charges for water treatment by this technology would be $24.30/month. The technology difference results in a cost increase of 80 percent. A similar example can be applied for drinking water treatment technologies.

As another example, assume flue gas desulfurization is used on a 500 MWe electrical generating station. A typical household receiving electricity from this plant uses 500 kWh/month. At an assumed electricity price of $0.05/kWh, the total monthly bill is $25.00/month. From Figure 5-8 in this report (Section 5), the annualized cost per kWh (unit annualized cost) for limestone flue gas desulfurization on a 500 MWe steam electric generating plant is $0.014/kWh. The impact on the typical monthly electrical bill for these conditions would therefore be about $7.00/month.

Where an environmental control technology is used in a manufacturing establishment, the relationship between the cost of control and consumer prices is more difficult to define. Examples of such technologies are industrial wastewater treatment, particulate matter control, and possibly flue gas desulfurization. If data on the manufacturing cost per unit of consumer product and the quantity of pollutant stream generated per unit of product were known for any specific article or industry, the calculation of the cost impact of the control technology on the consumer price would be straightforward. An industry-by-industry analysis is, however, clearly beyond the scope of this report. But, an approximate average relationship between control costs during production of a particular industrial product and consumer expenditures for that product can be derived for illustration. An example for particulate matter control applied to a consumer products industry, a building-material plant,† is discussed below.

A typical plant might produce about 400 million sq ft/yr of product. It would produce 40,000 acfm of particle-laden gas (air) requiring treatment with an electrostatic precipitator to remove particulate matter. About 1700 acf of gas would be treated for each standard unit (32 sq ft) of product produced. Referring to Figure 4-9 (Section 4), the unit annualized cost for an electrostatic precipitator with 99.9 percent removal and typical precipitation characteristics treating 40,000 acfm is $0.021/1000 acf. Multiplying this cost by 1700 acf/standard unit of product yields a cost of control per standard unit of product of cents per unit. If the product sells for about $3.40 per standard unit, the particulate matter control technology adds about 1.1 percent to the price the consumer pays in this example. The same concept can be applied to any other manufacturing industry, for any environmental control technology.

These illustrations are given only to provide perspective on the magnitude of impacts that environmental control technologies may have, and are only approximations. A detailed analysis is beyond the scope of this report. The examples are, however, an indication of how the environmental technology control costs can be reflected in consumer prices, and how information in this report can be used in estimating effects on prices.

---

*This is a rough estimate for a household of three people. In this report, system design capacity assumed a design value of 150 gallons per capita per day for wastewater and 200 gallons per capita per day for drinking water. Actual usage in a given household will not necessarily reflect these design values.

†Product details are not given so as to avoid any chance of misrepresentation of environmental control cost impacts for a specific product. A more detailed analysis would be needed to confirm production data, prices, and cost impacts for the actual industry.

References - Section 1

1. Uhl, V.W. A Standard Procedure for Cost Analysis of Pollution Control Operations. Vol. I. User Guide, EPA-600/8-79-018a (PB80-108038*). Vol. II. Appendices, EPA-600/8-79-018b (PB80-108046*). U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Research Triangle Park, NC, June 1979.
2. U.S. Department of Energy. Monthly Energy Review. DOE/EIA 0035/05(80).
3. U.S. Department of Labor. Monthly Labor Review., 103(6): June 1980.
4. Dames and Moore. Construction Costs for Municipal Wastewater Treatment Plants: 1973-1977. EPA-430/9-77-013 (PB282436*). U.S. Environmental Protection Agency, Office of Water Program Operations, January 1978.
5. Patterson, W.L., and R.F. Banker. Estimating Costs and Manpower Requirements for Conventional Wastewater Treatment Facilities. Project No. 17090 DAN, Contract No. 14-12-462, PB211132*. U.S. Environmental Protection Agency, October 1971.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

Section 2
Drinking Water Treatment

In 1975, some 177 million people were served by the approximately 40,000 community water systems in the United States (1). In addition to community water systems, numerous individual systems exist including individual households, and systems such as those at resorts or other public-access facilities. Most water systems serve a population of 100 to 10,000 people (1). Assuming a system design basis of 200 gallons per day (gpd) per capita, these systems are in the size range of about 20,000 gpd to 2.0 million gallons per day (mgd).

The raw water treated in these systems is either surface water from lakes and rivers or underground water. The purpose of these treatment systems is to make the water palatable, in terms of clarity, taste, and odor, and safe for human consumption. Treatment methods vary according to the quality of each individual water source.

Some contaminants occur naturally, some arise from domestic, industrial and agricultural activities, and some are formed during traditional water treatment. For example, trihalomethanes can form during conventional chlorine disinfection of drinking water when chlorine reacts with some organic substances.

The ionic species and organic compounds of interest in drinking water are usually expressed in concentration units of milligrams per liter (mg/l) or micrograms per liter (μg/l). Turbidity, caused by suspended solids, is usually expressed in turbidity units (TU) which are defined for a specific turbidity test method.

New knowledge of health effects and increasingly sensitive analytical chemistry procedures have enhanced recognition of potential long term health hazards due to certain water contaminants. This consideration, combined with increasing demands of population growth on available water supplies, requires that continued attention be given to drinking water treatment for upgrading raw water quality. The continued increase in demand relative to supply will likely increase the use of treated and recycled wastewater to meet drinking water needs in the future. More sophisticated methods and extensive use of these methods for drinking water treatment will be required.

A key legislative milestone was the Safe Drinking Water Act of 1974 (Public Law 93-523) and the promulgation of Interim Primary Drinking Water Regulations under that Act. This act focused the attention of the public and health and environmental professionals on the quality of drinking water supplies and resulted in drinking water quality standards to protect the consumer. The Act defines contaminants, maximum concentration levels, primary drinking water regulations, secondary drinking water regulations, public water supplies and systems, and other items. A set of enforceable health-related regulations and a set of non-enforceable aesthetic-related guidelines for drinking water were established. These regulations and subsequent revisions in 1978 set maximum levels for various water contaminants including potentially toxic ionic species, certain organic chemical compounds, and suspended solids which cause turbidity. Other materials that must continue to be removed include pathogenic microorganisms and substances which cause taste, odor, and color. Table 2-1 lists the permissible concentrations of various materials as set forth under the Act (2).

These regulations, which require greater removal of contaminants than is now common practice, increased the costs of treatment. The capital investment and annual costs of those treatment technologies required to meet the new standards are summarized in this section.

Because the cost of a water treatment system per unit of water produced decreases as plant size increases, the economic impact of increased water treatment on small systems is greater on a unit basis than on large systems. However, the total sums required for capital investment and operating requirements become large as system sizes increase. Policy planning must therefore address the cost implications of both large unit revenue requirements (higher customer costs) for small systems and investment capital availability for large systems. Finally, the cost impacts resulting from new treatment requirements must be viewed in the context of the total costs of the water supply system.

A charge rate profile for typical water supply systems was presented by Clark and Stevie (3). Table 2-2 presents the percentage contribution of each component of the overall water supply system, as derived from the data given by Clark. The average

Table 2-1. Drinking Water Contaminant Levels Based on Safe Drinking Water Act (2)

Contaminant	Maximum contaminant level (MCL)
Arsenic, mg/l	0.05
Barium, mg/l	1.0
Cadmium, mg/l	0.01
Chromium, mg/l	0.05
Lead, mg/l	0.05
Mercury, mg/l	0.002
Nitrate (as N), mg/l	10.0
Selenium, mg/l	0.01
Silver, mg/l	0.05
Endrin mg/l	0.002
Lindane mg/l	0.004
Toxaphene mg/l	0.005
2,4-D, mg/l	0.1
2, 4, 5 - TP (Silvex), mg/l	0.01
Methoxychlor, mg/l	0.1
Alpha emitters:	
Radium - 226, pCi/l	5
Radium - 228, pCi/l	5
Gross alpha activity (excluding radon and uranium), pCi/l	15
Beta and photon emitters:[a]	
Tritium, pCi/l	20
Strontium, pCi/l	8
Turbidity, turbidity unit[b]	1
Fluoride, mg/l[c]	1.4-2.4
Trihalomethanes and organic chemicals,[d]	

[a] Based on a water intake of 2 liters/day. If gross beta particle activity exceeds 50 pCi/l, other nuclides should be identified and quantified on the basis of a 2-liter/day intake.

[b] One turbidity unit based on a monthly average. Up to 5 turbidity units may be allowed for the monthly average if it can be demonstrated that no interference occurs with disinfection or microbiological determinations.

[c] Depends on air temperature.

[d] On February 9, 1978, the EPA proposed to amend the National Interim Primary Drinking Water Regulations by adding regulations for organic chemical contaminants in drinking water. The proposed amendment consisted of two parts:

1. An MCL of 0.10 mg/l (100 parts per billion) for total trihalomethanes (TTHM), including chloroform.
2. A treatment technique recommending the use of granular activated carbon for the control of synthetic organic chemicals. Three criteria that the granular activated carbon must achieve are: an effluent limitation of 0.5 $\mu g/l$ for low molecular weight halogenated organics (excluding trihalomethanes), a limit of 0.5 mg/l for effluent total organic carbon concentration when fresh activated carbon is used, and the removal of at least 50 percent influent total organic carbon when fresh activated carbon is used.

Part 1 was promulgated November 29, 1979. Part 2 has been cancelled.

Table 2-2. Charge Rate Profile for Typical Total Drinking Water System Costs Based on Survey Data (3)

	Percentage of total	
System component	Large system (85 mgd)	Small system (5 mgd)
Support services	24.4	17.6
Acquisition	13.4	15.1
Treatment	11.8	10.3
Distribution	29.0	41.9
Interest on debt	21.4	15.1

large system capacity was 85 mgd, and the average small system was 5 mgd. If the costs reflected in the charge rate presented by Clark are updated to March 1980, the typical total system charge rate is \$0.57/1000 gal for large systems and \$1.20/1000 gal for small systems. If it is assumed that user rates are approximately 20 percent greater than the system charge rates, these figures provide an estimate of typical user total charge rates.

It must be emphasized that these costs are for existing systems rather than new facilities; therefore, costs included for capital-related charges are based on historical values for invested capital. Because the water systems in the study were built 30 to 40 years ago, the capital charge components in the total are far lower than would be encountered if comparable facilities were built today. On a historical basis, the cost of the treatment step is approximately 11 percent of the total cost of supplying drinking water. In considering upgrading water supply systems to improve water quality, however, a main focus of the upgrading will be on the treatment technologies.

Where an acquisition and distribution system is already in place, a new treatment facility might be built to upgrade or replace the existing treatment facility. In such a case, only the treatment portion of the water cost would have to be substantially changed. Obviously, its portion of the total system cost would increase from the 11 percent discussed above.

Based on the considerations discussed above, only treatment technologies are considered in this report. For some water systems, these treatment technologies might be necessary to meet water quality requirements under the Safe Drinking Water Act. These technologies include both total treatment systems as well as individual single processes which could be added at existing total treatment plants or incorporated into new total treatment plants. The single processes include methods for disinfection and the removal of organic chemical compounds.

The total treatment systems comprising several process steps include conventional filtration, direct filtration, and lime softening filtration plants. These filtration plants primarily reduce the concentration of both dissolved inorganic materials and suspended solids present in the raw water. Most of the toxic substances identified in the Safe Drinking Water Act are probably present as dissolved solids. Suspended solids cause turbidity and harbor harmful microbes.

Direct filtration differs from conventional filtration primarily by the absence of the sedimentation step. Chemicals such as alum or iron salts are added to precipitate suspended solids directly in the filters. In lime softening plants the addition of lime contributes not only to suspended solids removal but also to the removal of some dissolved substances. These dissolved substances become insoluble upon the addition of lime and are removed by sedimentation and filtration. Lime is used when lower levels of dissolved minerals must be achieved than can be achieved using alum or iron salts.

Disinfection methods include chlorination and treatment with other disinfectants. In some water systems where raw water quality is high, chlorine disinfection may be the sole treatment used. Alternative disinfectants to chlorine include chlorine dioxide, ozone, and ammonia used in combination with chlorine. Granular activated carbon treatment and aeration are technologies used specifically for the removal of organic compounds. Disinfection and organic removal methods can be add-on technologies to any of the total treatment processes.

In the individual technology sections that follow, cost curves are presented for total capital investment, net annual operating expenses, and unit annualized cost as a function of system design capacity in millions of gallons per day. Brief process descriptions, a design basis for the costs presented, and a summary of major variables affecting costs are also presented.

2.1 Filtration Treatment Plants

2.1.1 Description

There are three different kinds of filtration treatment plants:

- Conventional filtration.
- Direct filtration.
- Lime softening (with conventional filtration).

These three kinds of plants share several common process steps. Direct filtration and lime softening are essentially variations of a conventional filtration plant. These plants all remove turbidity-causing suspended solids and some mineral matter from drinking water supplies.

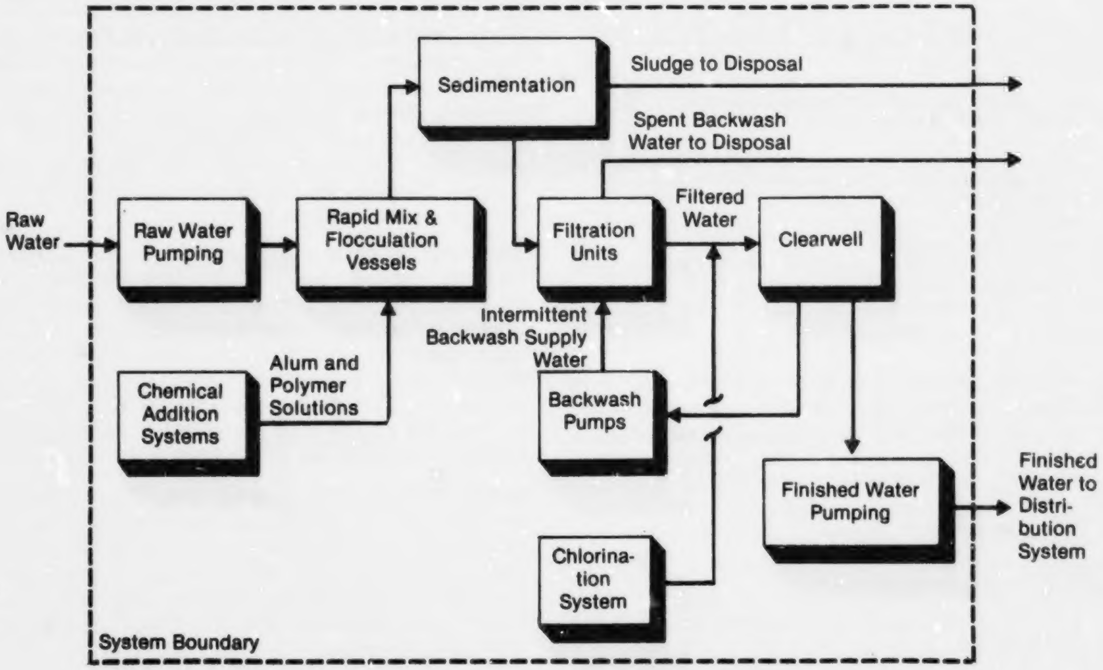
Conventional Filtration (2,4,5)

A conventional filtration plant for drinking water treatment removes suspended solids and some dissolved mineral matter. It also destroys harmful microorganisms in the water supply. A typical conventional filtration plant is shown conceptually in Figure 2-1.

Raw water is pumped to a rapid mix tank in which chemical (e.g., alum and polymer) solutions are added to enhance flocculation. A flocculation vessel allows sufficient time for the suspended solids to aggregate into the larger particles, or flocs, which are more efficiently removed by the downstream treatment steps (sedimentation and filtration).

Sedimentation basins are either circular or rectangular vessels in which the flocs are allowed to settle. The basins can be concrete or steel, depending on size. A waste sludge of solids and water is removed by

Figure 2-1. Conventional filtration system for drinking water treatment.


Sedimentation
Sludge to Disposal
Spent Backwash Water to Disposal
Raw Water
Raw Water Pumping
Rapid Mix & Flocculation Vessels
Filtration Units
Filtered Water
Clearwell
Intermittent Backwash Supply Water
Alum and Polymer Solutions
Chemical Addition Systems
Backwash Pumps
Finished Water Pumping
Finished Water to Distribution System
Chlorination System
System Boundary

discharge to a municipal sewer or hauled to a landfill for disposal. Clarified water then flows to the filter unit.

The filters consist of one or more steel or concrete vessels containing granular materials such as graded sands, anthracite, and garnet. Solids are strained from the water as it passes through the filters. When the pressure drop through the filters becomes great enough due to accumulated solids, a backwash stream of filtered water passes through the units in reverse flow to clean the solids from the filter bed. The spent backwash stream is sent to a sewer. Backwashing is intermittent; the backwash cycle depends on the character and concentration of solids in the water, as well as on filter design parameters such as application rate and filter medium particle size.

Filtered water is disinfected with chlorine and stored. From storage it is pumped to the water supply distribution system.

Direct Filtration (2,4,5)

A direct filtration plant is essentially the same as the conventional filtration plant shown in Figure 2-1 except the sedimentation step is deleted.

Direct filtration is applicable to any drinking water supply where suspended solids levels are sufficiently low to result in a reasonable backwash cycle on the filter units. Unlike conventional filtration plants, there is an upper limit to the influent suspended solids concentration that can be tolerated. This upper limit must be determined by testing. Above such a level, conventional treatment procedures or sedimentation prior to filtration are required.

Lime Softening (2,4,5)

The major features of a lime softening plant are also essentially the same as those for a conventional filtration plant, except that lime is substituted for other chemicals and a recarbonation step is added after sedimentation. A lime softening plant is typically used to treat raw water with a higher concentration of dissolved minerals, such as calcium and magnesium, than can be treated in a conventional or direct filtration plant. In the context of the Safe Drinking Water Act, a lime softening plant can also be expected to achieve a greater removal of toxic mineral substances. For example, a lime softening plant operating in a pH range of 8.5 to 11 can reduce cadmium concentrations from 0.5 mg/l to 0.01 mg/l. To achieve the same cadmium concentration in the treated effluent, a conventional filtration plant using alum or iron salts can only accommodate a cadmium concentration up to 0.1 mg/l of cadmium in the raw water (2). The choice of overall treatment process therefore depends on individual raw water characteristics.

Lime can be added directly to the influent raw water as a solid, or as a pre-mixed water slurry. If a slurry is used, the solid lime is usually purchased and the slurry prepared on-site. Details of lime feed systems are described elsewhere (6, 7).

Recarbonation is the addition of gaseous carbon dioxide (CO_2) to the lime-treated water to neutralize excess alkalinity resulting from lime addition. Gaseous CO_2 may be obtained from liquid CO_2 stored onsite, submerged burners, or stack gas compressed through a sparger system. The choice of carbonation method depends on site specific considerations.

2.1.2 Design Basis and Costs (2,4,5)

The design basis in this report for conventional filtration plant costs includes the following major process modules and design parameters:

- Raw water pumping.
- Chemical addition.
- Rapid mix/Flocculation.
- Sedimentation.
- Filtration.
- Disinfection.
- Finished water storage.
- Finished water pumping.
- Sludge disposal.

As stated in the process descriptions, there is no sedimentation step in direct filtration. The filtration directly follows the rapid mix and flocculation step. The chemical feed system consists of chemical storage and metering pump facilities. The rapid mix tank and flocculation vessel is one vessel partitioned into separate sections. Filtration units are gravity flow steel or concrete vessels. The clear well is a concrete storage basin. System design parameters depend on raw water quality and the finished water quality required.

The major process modules for the lime softening plant are very similar to those for conventional filtration, except for modifications to the chemical feed system and addition of recarbonation equipment. Recarbonation basins are reinforced concrete, and submerged natural gas burners are used for the CO_2 source in the system considered here based on the configuration and costs in Reference 2.

The plant cases represented here include chlorine disinfection, the usual procedure in conventional plants. Alternative disinfectants such as chlorine dioxide, ozone, or ammonia added with chlorine can also be used. The disinfection systems for each of these alternatives are discussed in Section 2.2

Total capital investment for conventional filtration, direct filtration, and lime softening is presented in Figure 2-2. Net annual operating expenses are shown in Figure 2-3. Figure 2-4 shows corresponding unit annualized costs.

Figure 2-2. Filtration plants for drinking water treatment - Total capital investment (March, 1980 dollars).

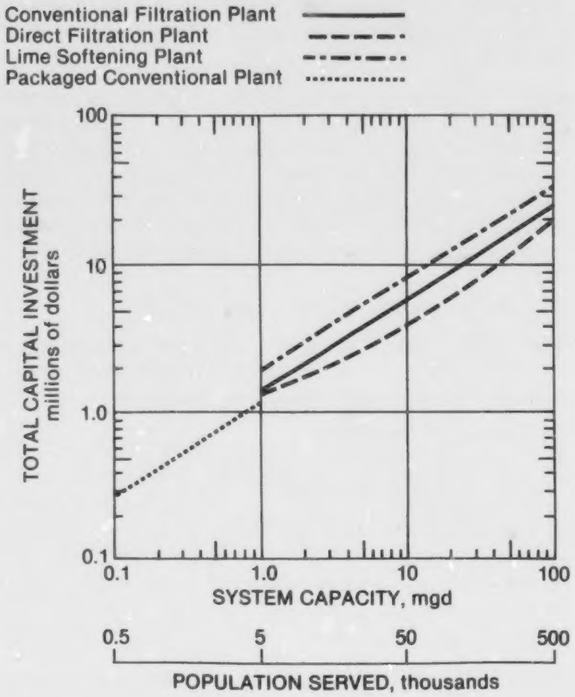


Figure 2-3. Filtration plants for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

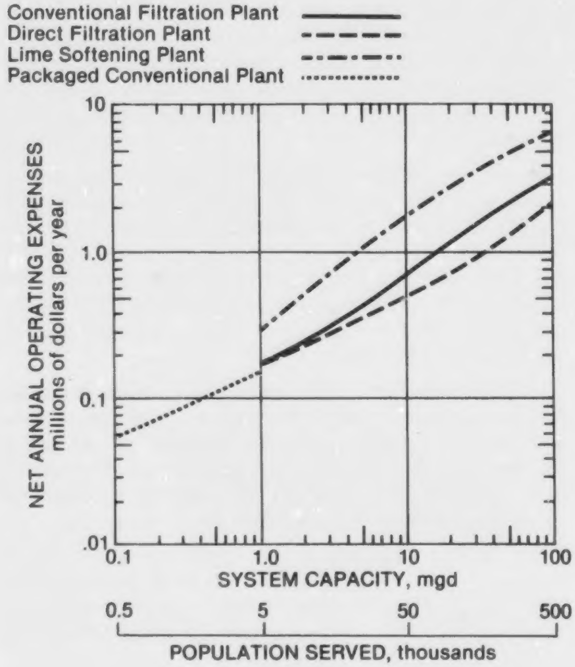
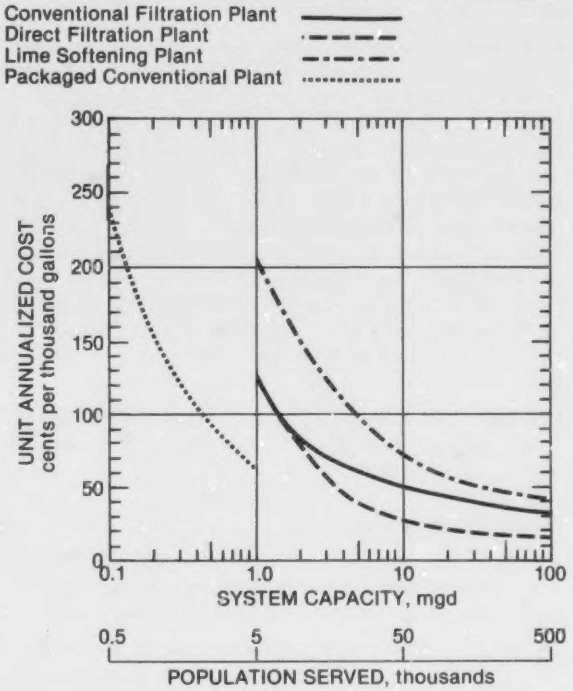


Figure 2-4. Filtration plants for drinking water treatment - Unit annualized cost (March, 1980 dollars).

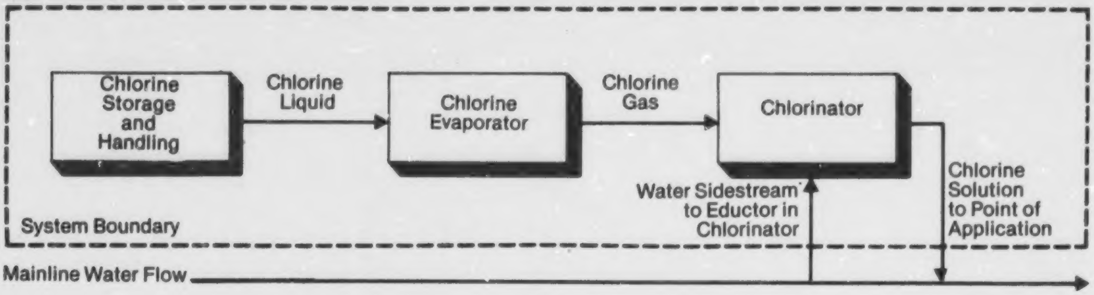


Also provided in the figures are costs for packaged conventional filtration plants which can be used for small treatment systems (5). These plants would have the same unit processes as their larger field-constructed counterparts but would be primarily shop fabricated and brought to the field for final installation.

2.1.3 Major Variables Affecting Costs

For any of the filtration plants discussed here, the large number of process steps and associated variables result in many possible combinations of equipment sizes and specifications. These factors largely depend on site specific requirements with raw water quality the primary variable. A complete analysis of the cost impacts of changes in design is beyond the scope of this report. However, examination of the cost profile for capital investment reveals that the greatest portion of the investment is in the filter portion of the plant. Therefore, changes in design requirements for the filters have a very large impact on total plant capital costs. For lime softening plants lime dosage is an important variable. Also, as can be seen from the figures, costs for shop fabricated packaged plants are less than for field constructed plants of similar size. Operating expenses, specifically electricity costs for pumping, are affected by frequency of backwashing in the filtration unit which

Figure 2-5. Chlorination system for drinking water treatment.


Chlorine Storage and Handling
Chlorine Liquid
Chlorine Evaporator
Chlorine Gas
Chlorinator
Chlorine Solution to Point of Application
Water Sidestream to Eductor in Chlorinator
System Boundary
Mainline Water Flow

in turn depends on raw water suspended solids levels.

2.2 Disinfection

Disinfection destroys microbes harmful to human health. Chlorine is the most commonly used disinfectant. Because chlorine addition can lead to the formation of trihalomethanes (potential carcinogens), use of the alternative disinfectants chlorine dioxide, ozone, and ammonia in combination with chlorine has been considered (8). The last alternative results in the formation of chloramines which disinfect while obviating the formation of trihalomethanes.

2.2.1 Chlorine

Description (4)

Chlorine may be added directly to the water as chlorine gas or indirectly as a sodium hypochlorite solution. Only direct feed chlorination is discussed in this report because it is more widely used.

The major features of a chlorination system are shown in Figure 2-5. The system includes both chlorine storage and feed equipment. The chlorinator consists of a metering device and an educator in which the chlorine mixes with a small sidestream taken from the main water line. After passing through the chlorinator, the sidestream rejoins the main flow, delivering the disinfectant to the water supply.

For small systems that require chlorine feed rates at 100 lb/day or less, chlorine is stored in standard 150-lb cylinders. Chlorine for larger systems with feed rates up to 2000 lb/day is stored in 1-ton cylinders. For systems larger than 2000 lb/day, chlorine is stored in: 1-ton cylinders, on-site tanks supplied by rail delivery, or rail cars kept on a rail siding.

Design Basis and Costs (2,4,5)

Each installation is assumed to have a duplicate standby chlorinator, injector pumps on the water sidestream fed to the chlorinator, housing for the chlorinator, and a 30-day chlorine storage capacity. Cylinder storage is assumed. Evaporators are assumed for systems requiring chlorination feed rates of greater than 2000 lb/day; chlorine residual analyzers are assumed for systems where chlorine flow rates are greater than 1000 lb/day; cylinder hoists are assumed for systems where chlorine feed rates are less than 100 lb/day.

Typical piping costs are included, although individual site layouts will cause these to vary.

Operating requirements include labor for operation and maintenance of the metering equipment, and activities associated with storage. Material requirements are for maintenance. Power requirements are for pumping, mixing, and building heating, lighting, and ventilation.

Total capital investment requirements for chlorination systems are presented in Figure 2-6. Total capital investment is plotted against water plant flow rate in millions of gallons per day. The two curves correspond to chlorine feed dosage rates of 1 mg/l and 5 mg/l. The dosage rate required depends on the disinfection requirements of the specific water supply being treated. Below plant capacities of 0.1 mgd, the total capital investments is essentially at about $7,600 (March 1980 dollars) (5).

Net annual operating expenses are plotted against water plant flow rate in millions of gallons per day in Figure 2-7. Unit annualized cost is shown in Figure 2-8. The two curves again reflect different chlorine dosage levels.

Major Variables Affecting Costs

The cost curves indicate clearly the effect of dosage on costs for chlorination systems. Dosage in turn depends on individual water characteristics. An important variable in chlorination is pH as it affects the chemistry of solution and hence the dosage required to achieve a given disinfection effectiveness (8). The type of storage system in larger facilities, tank or railcar siding storage compared to the cylinder storage used here, for example, and individual plant layout items such as differing lengths of piping runs also affect costs.

Figure 2-6. Chlorination system for drinking water treatment - Total capital investment (March, 1980 dollars).

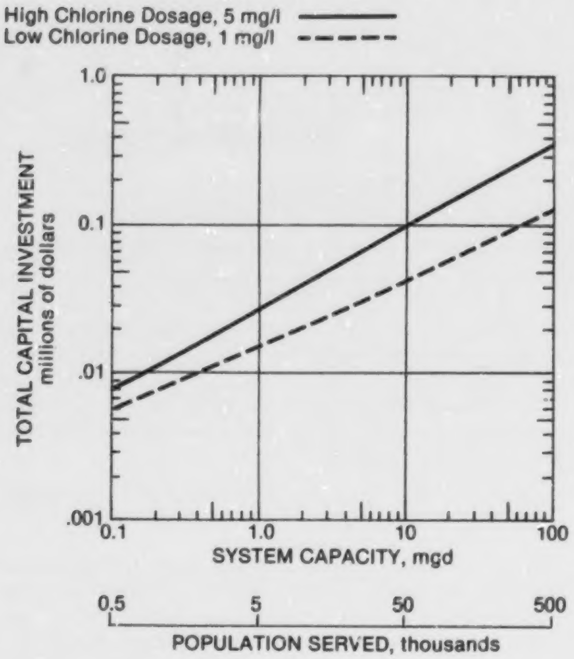


Figure 2-7. Chlorination system for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

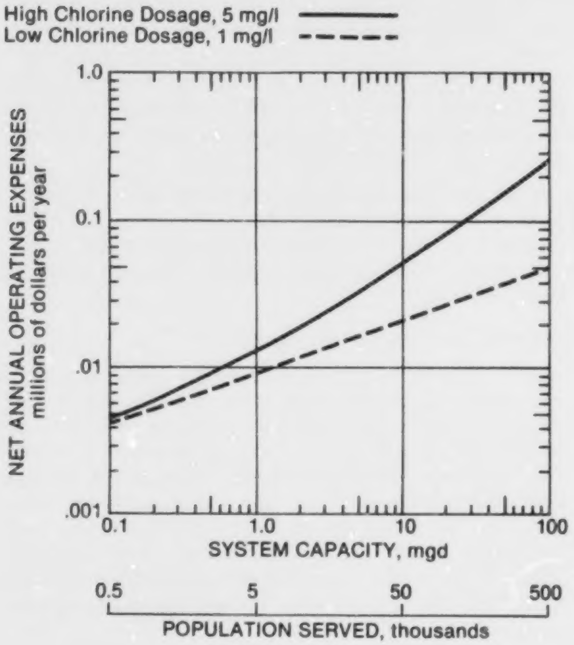
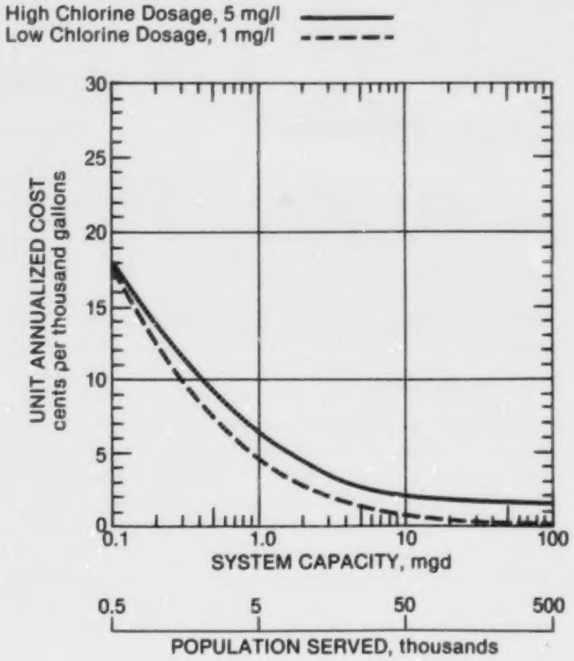


Figure 2-8. Chlorination system for drinking water treatment - Unit annualized cost (March, 1980 dollars).



2.2.2 Chlorine Dioxide

Description (4)

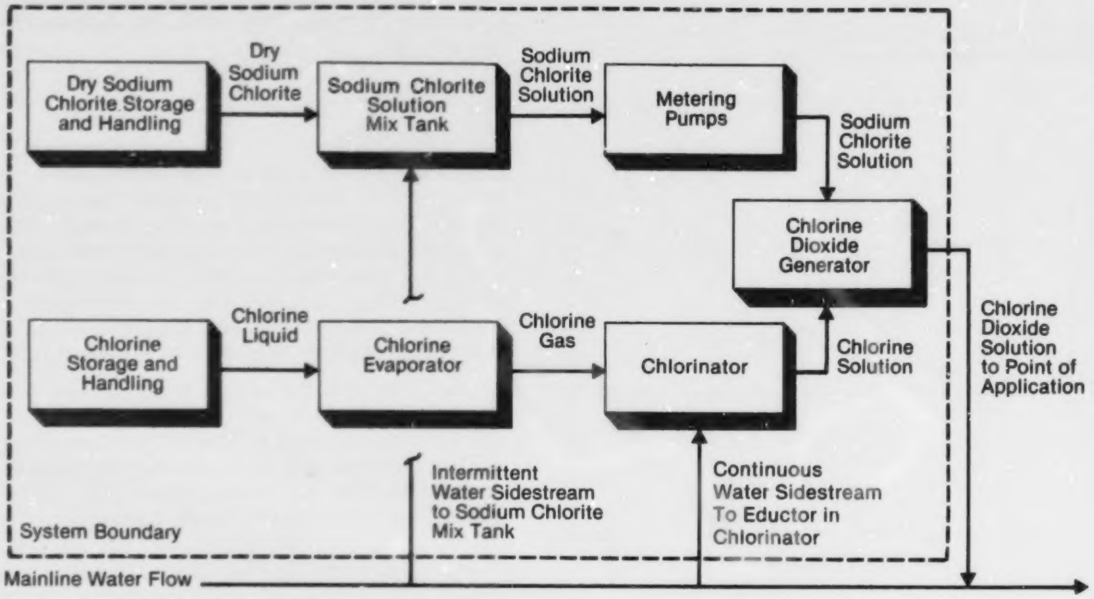
Chlorine dioxide is used for disinfection of drinking water in the same manner as chlorine. In fact, the feed equipment for chlorine dioxide is essentially the same as that for chlorine. Rather than obtaining chlorine dioxide from storage containers, as is done with chlorine, however, chlorine dioxide gas is commonly generated on-site by mixing a high-strength chlorine solution with a high-strength acidified sodium chlorite solution. These solutions are fed to a mixing chamber referred to as a generator. The generator is a plastic cylinder containing a loose porcelain fill material. Detention time in the generator is about 2 minutes or less. The gas evolving from solution then feeds to a device identical to the chlorinator discussed for chlorine treatment in Section 2.2.1.

A typical schematic of a chlorine dioxide system is shown in Figure 2-9. The sodium chlorite system consists of a polyethylene mix tank and a metering pump. The sodium chlorite is stored in bags on pallets.

Design Basis and Costs (2,4,5)

To generate 1 lb of chlorine dioxide, a feed ratio of 1.68 lb chlorine to 1.68 lb sodium chlorite is assumed.

Figure 2-9. Chlorine dioxide system for drinking water treatment.


Dry Sodium Chlorite Storage and Handling
Dry Sodium Chlorite
Sodium Chlorite Solution Mix Tank
Sodium Chlorite Solution
Metering Pumps
Sodium Chlorite Solution
Chlorine Dioxide Generator
Chlorine Dioxide Solution to Point of Application
Chlorine Storage and Handling
Chlorine Liquid
Chlorine Evaporator
Chlorine Gas
Chlorinator
Chlorine Solution
Intermittent Water Sidestream to Sodium Chlorite Mix Tank
Continuous Water Sidestream To Eductor in Chlorinator
System Boundary
Mainline Water Flow

In order to estimate costs the necessary equipment has been added to the design scope of the chlorine systems already discussed in the preceding section. Costs for bag storage of sodium chlorite on pallets are assumed to be negligible in the context of the total system and were not included in the costs in References 2, 4, and 5.

Operating requirements include labor for the chlorine system as well as labor for preparation of the hypochlorite solution and for maintenance of mixing and metering equipment. Material requirements are for maintenance of all system components. Power requirements are for pumping, mixing, and building heating, lighting, and ventilation.

Capital costs for chlorine dioxide systems are presented in Figure 2-10 expressed as total capital investment plotted against water system flow rate in millions of gallons per day. The two curves correspond to chlorine dioxide dosage rates of 1 mg/l and 5 mg/l. The dosage rate required depends on the disinfection requirements of the specific water supply being treated.

Figure 2-11 presents net annual operating expenses plotted against water system flow rate in millions of gallons per day. Again, the two curves correspond to different chlorine dioxide dosage levels. Unit annualized costs are shown in Figure 2-12.

Major Variables Affecting Costs

The cost curves indicate clearly the effect of dosage on costs for chlorine dioxide systems. Individual plant

Figure 2-10. Chlorine dioxide system for drinking water treatment - Total capital investment (March, 1980 dollars).

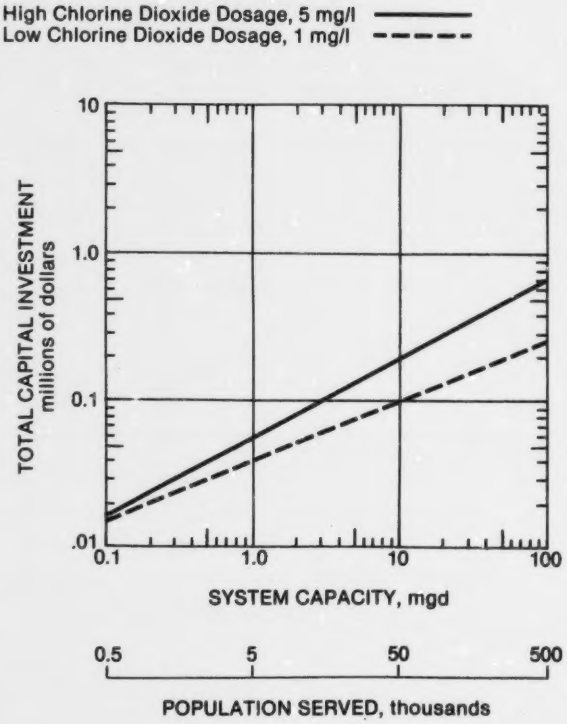


Figure 2-11. Chlorine dioxide system for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

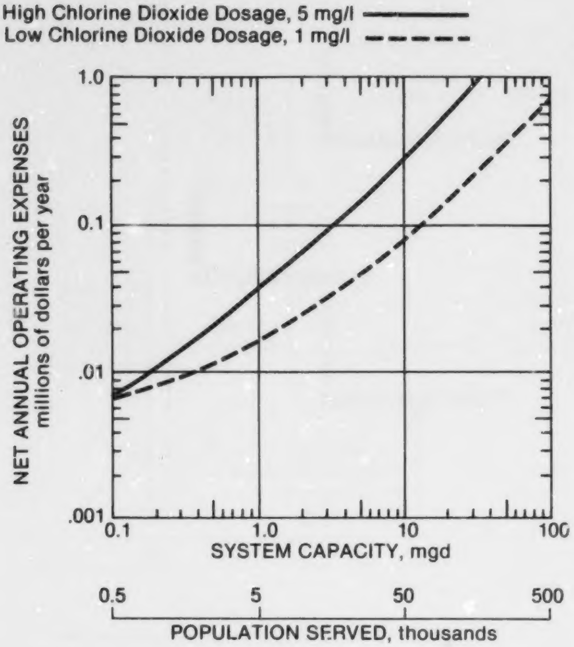
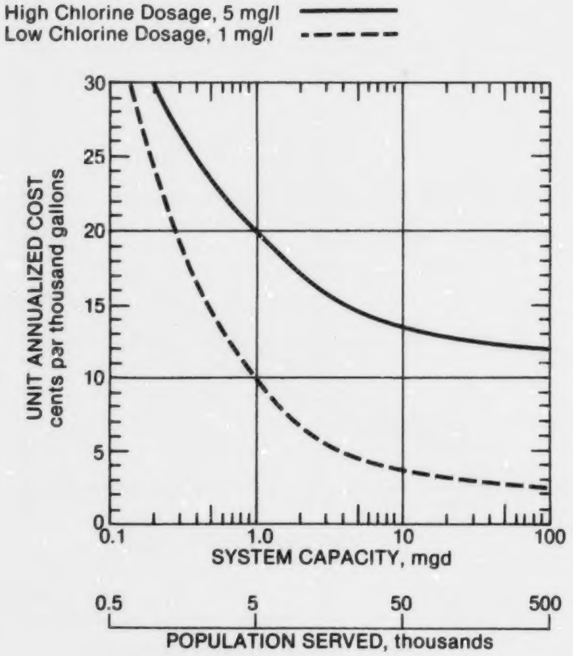


Figure 2-12. Chlorine dioxide system for drinking water treatment - Unit annualized cost (March, 1980 dollars).



layout and the storage system selected for the required chlorine starting material can affect costs significantly. Considerations similar to those discussed for chlorination systems apply for chlorine dioxide also.

2.2.3 Ozone

Description (4)

Ozone is generated on-site by passing air or oxygen through an electric arc. Ozone generators are standard items manufactured by vendors. At ozone generation rates less than approximately 100 lb/day air is more economical than oxygen. Pure oxygen storage costs can be justified at higher ozone generation rates. A block diagram of an ozone disinfection system is shown in Figure 2-13.

Ozone from the generator feeds to the dissolver chamber where it is well mixed with a sidestream of the water being treated. The solution then flows to a contact chamber where it mixes with the mainstream waterflow. The required contact time is typically about 15 minutes in the contact chamber.

Design Basis and Costs (4,5)

System costs include costs for the components shown in Figure 2-13, as well as the costs of equipment for off-gas recycling, electrical, instrumentation, safety, and monitoring requirements. The ozone contact chamber is a covered reinforced concrete structure, 18 ft deep with a length-to-width ratio of 2:1. The chamber contains partitions to ensure good flow distribution.

For systems that require 100 lb/day or less of ozone, air is the oxygen source. Systems with an ozone requirement greater than 100 lb/day use oxygen and include oxygen storage and transfer equipment. In the typical system, all equipment is housed except oxygen equipment which is located outside the building on a concrete slab.

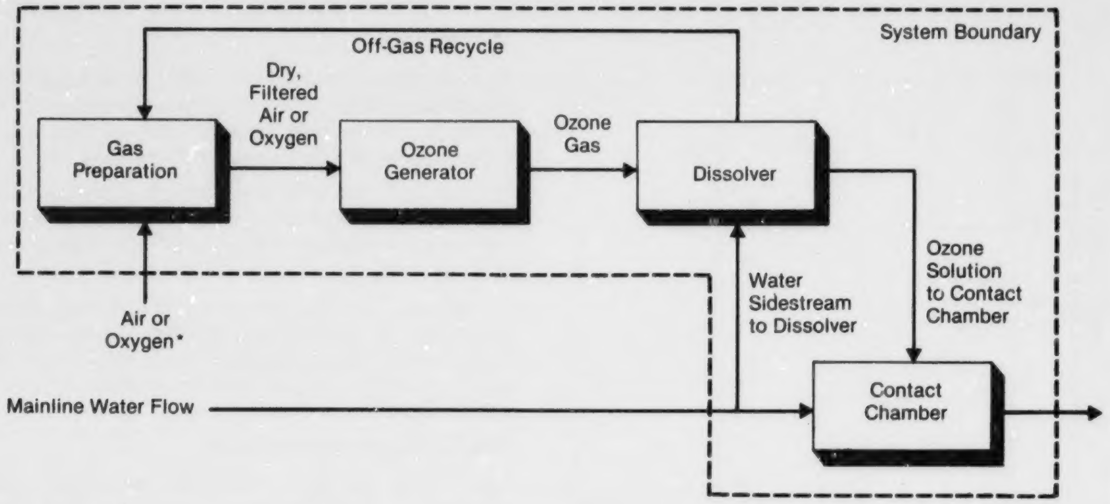
Operating requirements are similar to those discussed for chlorine and chlorine dioxide. Electrical power costs will be higher because electricity is used to generate ozone as well as for pumping and building requirements.

Total capital investment for ozone systems is shown in Figure 2-14, net annual operating expenses in Figure 2-15, and unit annualized cost in Figure 2-16. The two curves in each figure correspond to different dosage levels. The dosage level depends on the ozone demand of the specific water stream being treated.

Major Variables Affecting Costs

Dosage, which in turn depends on the characteristics of the individual water supply, has a pronounced effect on the cost of an ozonation system. It affects storage and feed equipment sizing, and electricity

Figure 2-13. Ozonation system for drinking water treatment.


System Boundary
Off-Gas Recycle
Dry, Filtered Air or Oxygen
Gas Preparation
Ozone Generator
Ozone Gas
Dissolver
Ozone Solution to Contact Chamber
Water Sidestream to Dissolver
Air or Oxygen*
Mainline Water Flow
Contact Chamber

*If oxygen is used the system will include oxygen storage equipment. Oxygen would be used only for ozone requirements greater than 100 lb/day.

Figure 2-14. Ozonation system for drinking water treatment - Total capital investment (March, 1980 dollars).

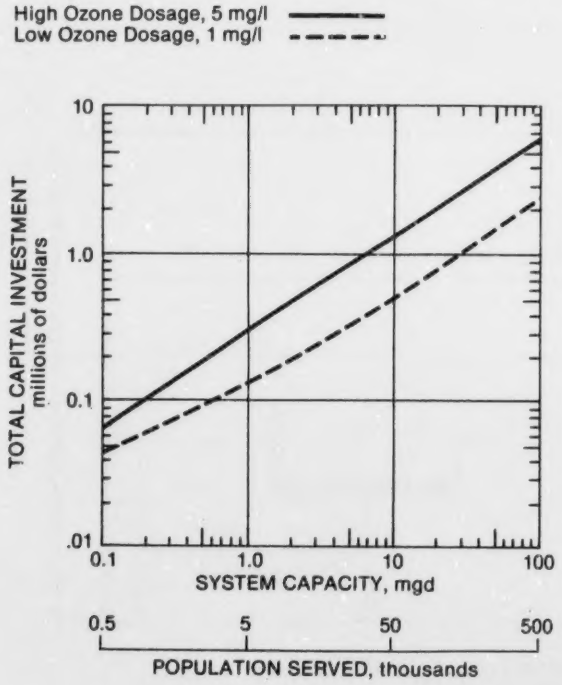


Figure 2-15. Ozonation system for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

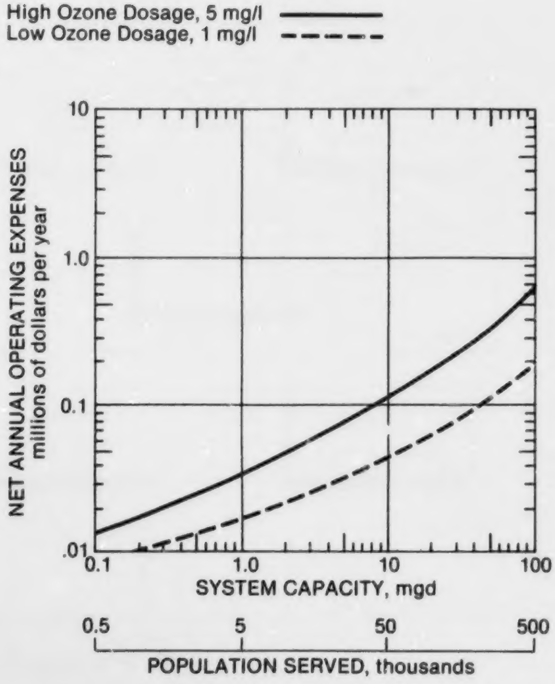
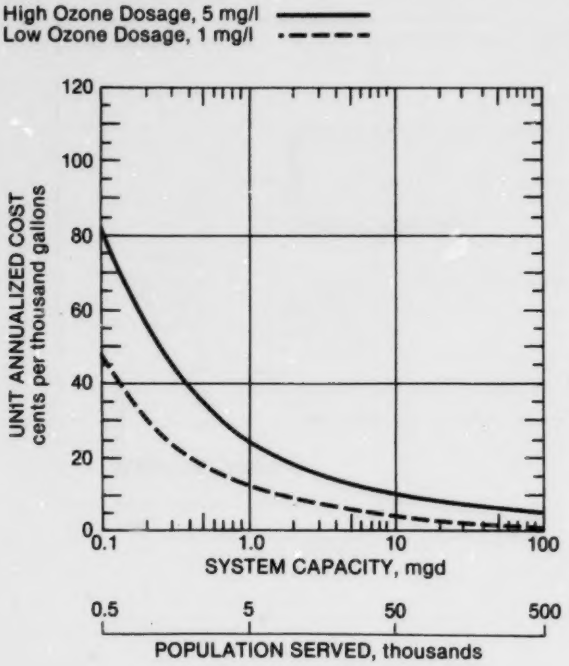


Figure 2-16. Ozonation system for drinking water treatment - Unit annualized cost (March, 1980 dollars).



requirements. System design, with the choice of either an air or oxygen feed, also influences costs.

2.2.4 Chloramination

Description

In chloramination, chlorine and ammonia are mixed together in water solution to form chloramines which act as a disinfectant. Chloramination does not form trihalomethanes as does direct chlorination.

Figure 2-17 is a schematic of a chloramination system. The system is comprised of a direct chlorination system with the addition of an ammonia feed system.

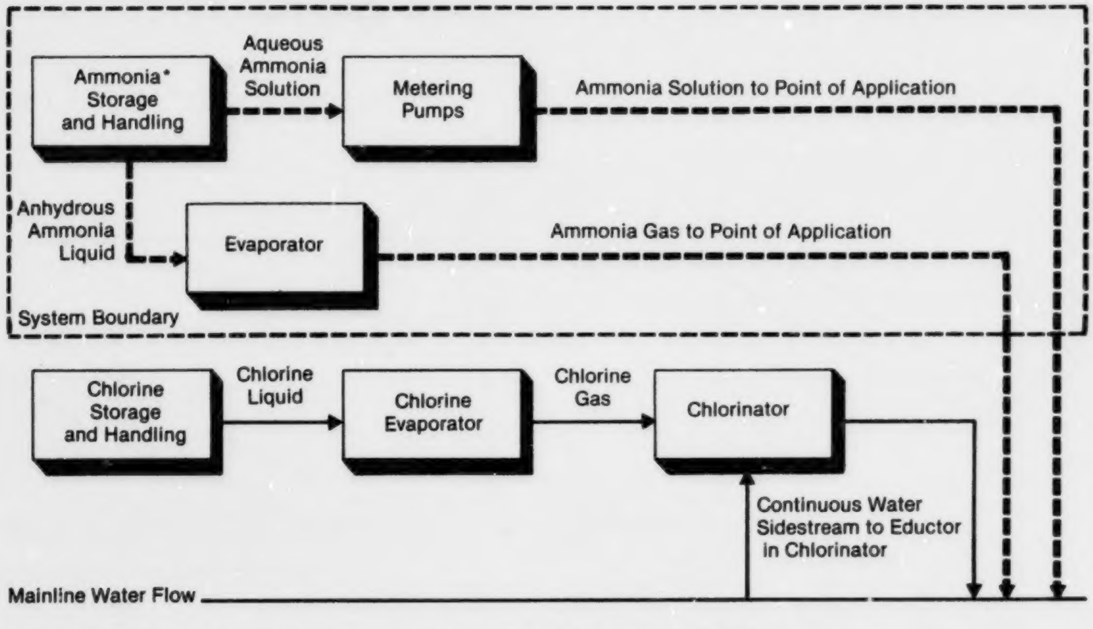
A system can be designed for either aqueous ammonia or anhydrous ammonia feed. Aqueous ammonia is usually available near large cities and is used more in larger facilities than anhydrous ammonia.

Design Basis and Costs (4)

The anhydrous ammonia system provides a 10-day storage capacity for bulk ammonia. The storage system includes the tank and its supports, a weigh scale, an air padding system for the tanks, and all gauges, pipes, and valves. The feed portion of the system consists of an evaporator and flow metering equipment.

The aqueous ammonia system also provides for a 10-day storage capacity. The storage system includes a

Figure 2-17. Chloramination system for drinking water treatment.


Ammonia*
Storage
and Handling
Aqueous
Ammonia
Solution
Metering
Pumps
Ammonia Solution to Point of Application
Anhydrous
Ammonia
Liquid
Evaporator
Ammonia Gas to Point of Application
System Boundary
Chlorine
Storage
and Handling
Chlorine
Liquid
Chlorine
Evaporator
Chlorine
Gas
Chlorinator
Continuous Water
Sidestream to Eductor
in Chlorinator
Mainline Water Flow

*One of the two ammonia systems shown would be selected for a given installation.

horizontal pressure vessel, supports, piping and valves, and a metering pump.

Costs are given only for the ammonia components of a chlorimination system. This technology would most likely be used to convert a plant with an existing chlorination system to chloramination. It is less likely that plants would be specifically designed to use chloramination. For those cases in which the ammonia and chlorine systems are constructed at the same time, however, the chlorination costs discussed earlier in Section 2.2.1 can be added to costs for the ammonia system to derive a total chloramination system cost.

Total capital investment is presented in Figure 2-18, net annual operating expenses in Figure 2-19, and unit annualized cost in Figure 2-20. The multiple curves shown correspond to different dosage requirements and the form of ammonia used. As the curves show, costs are relatively constant below a minimum system size.

Major Variables Affecting Costs

The most significant design variable which affects costs is whether anhydrous or aqueous ammonia is used. As mentioned earlier this can be partly influenced by geographical location and the relative availability of anhydrous and aqueous ammonia. Dosage is another significant variable which depends on individual water characteristics as it did with other disinfectants.

Figure 2-18. Ammonia feed system for drinking water treatment by chloramination · Total capital investment (March, 1980 dollars).

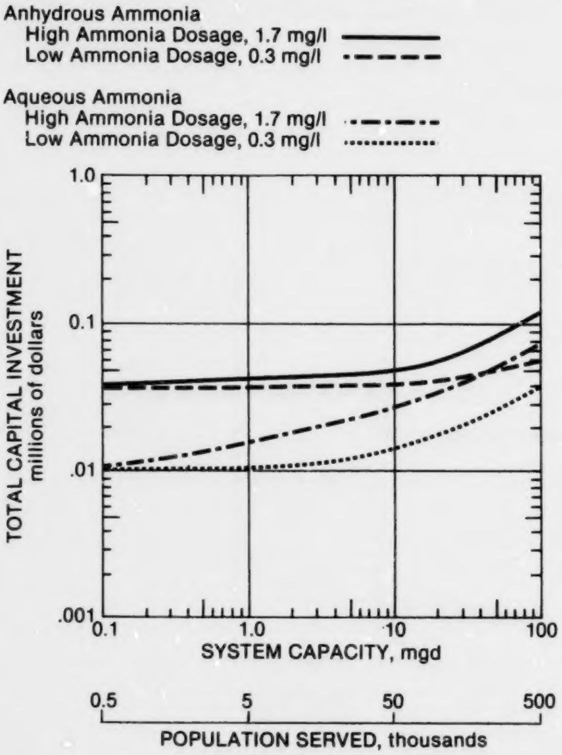
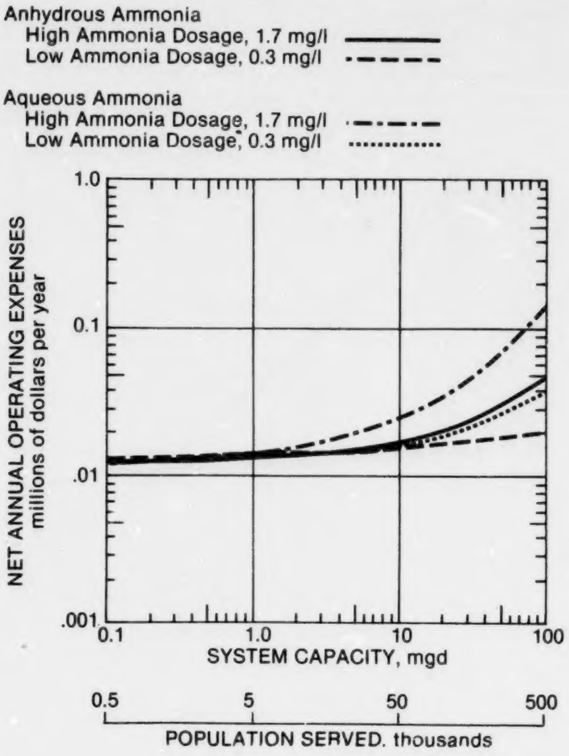


Figure 2-19. Ammonia feed system for drinking water treatment by chloramination · Net annual operating expenses (March, 1980 dollars).



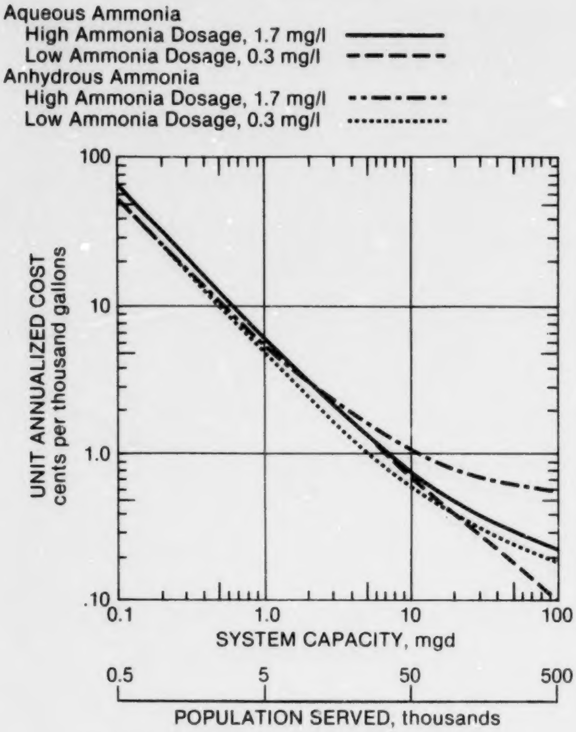
2.3 Granular Activated Carbon Treatment

Description

Granular activated carbon can be used in a drinking water treatment plant to remove dissolved organic compounds some of which may be present in low concentrations. These compounds may be present in the raw makeup water or they may be formed as a result of drinking water chlorination (e.g., trihalomethanes). Carbon treatment can be used either before or after chlorination to remove either precursors or contaminants themselves that might form.

A typical complete granular activated carbon treatment system is illustrated schematically in Figure 2-21. Water may enter the system after

Figure 2-20. Ammonia feed system for drinking water treatment by chloramination - Unit annualized cost (March, 1980 dollars).

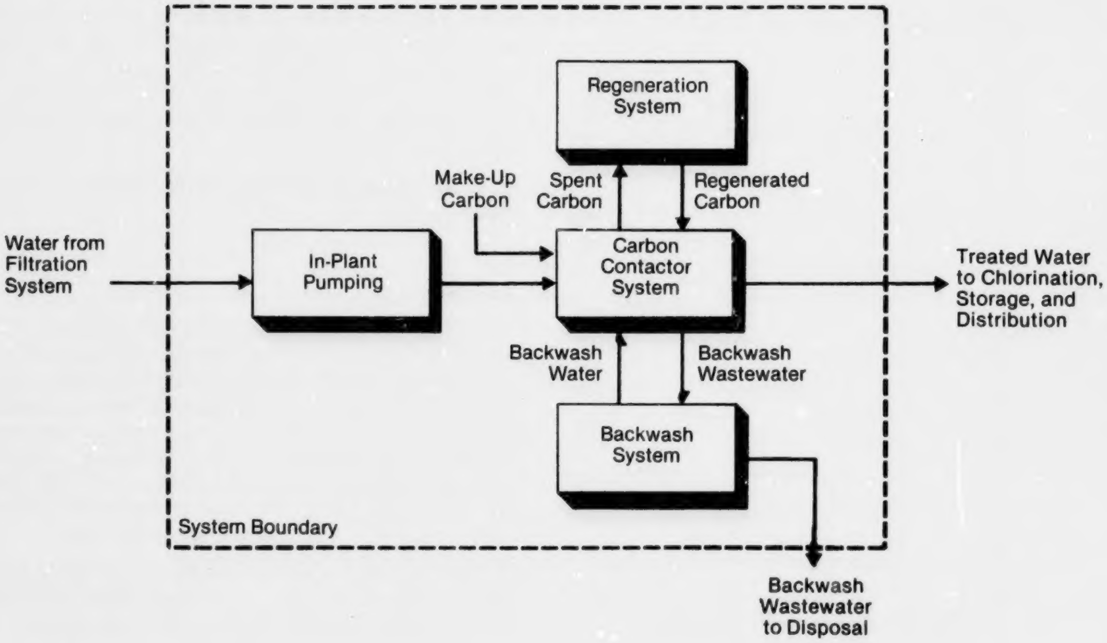


treatment to remove suspended solids and/or after lime softening used to remove mineral substances in one of the treatment plants already discussed in Section 2.1. It then flows by gravity or pressure through stationary beds of activated carbon contained in two or more steel or concrete adsorption vessels (contactors).

Packaged plants with pressure flow steel contactors may be used in small facilities with system capacities of less than 1 mgd. Plants treating more than 1 mgd are usually field constructed because mechanical, structural, and transportation constraints limit the size of portable units. These plants can use either pressure flow steel contactors or gravity flow concrete or steel contactors. In larger facilities above approximately 10 mgd, gravity flow concrete or steel contactors are generally used. Concrete contactors are usually more cost effective in large installations because large volume steel contactors are expensive, as are the large numbers of smaller steel vessels that would be required. Contactors are available in standard sizes, and multiple contactors operating in parallel flow are used to achieve a given plant capacity.

Periodically, typically every several months, the carbon must be removed from the contactors and regenerated to restore its ability to remove contaminants from water. This is accomplished by burning off the contaminants in a regeneration furnace. A granular activated carbon system contains at least two contactors so that they can be alternately

Figure 2-21. Granular activated carbon system for drinking water treatment.


Regeneration System
Make-Up Carbon
Spent Carbon
Regenerated Carbon
Water from Filtration System
In-Plant Pumping
Carbon Contactor System
Treated Water to Chlorination, Storage, and Distribution
Backwash Water
Backwash Wastewater
Backwash System
System Boundary
Backwash Wastewater to Disposal

regenerated. The carbon may be regenerated either on-site or in an off-site facility. Carbon regeneration facilities include multihearth, fluidized bed, or infrared furnaces. Small packaged plants may be designed for disposal of the spent carbon in lieu of regeneration.

Water treatment plants with sand bed filtration can be converted to granular activated carbon treatment by replacing the sand beds in the contactors with carbon and making other equipment modifications and additions.

Design Basis and Costs (2,4,5)

As discussed in the description of this technology, various kinds of contactors and regeneration methods are possible. Besides different kinds of equipment, the parameters for sizing the equipment will also determine characteristics for a given system.

Parameters such as hydraulic loading (application rate of water to the contactor cross-sectional area, expressed as gpm/ft^2), carbon depth, and regeneration frequency vary according to the kind and concentration of organics in the influent water, the final water purity required, and the type of carbon used. The design basis for costs presented here corresponds to the process modules shown in Figure 2-21. The major design criteria upon which costs are based are listed in Table 2-3. At system capacities of 1.0 mgd and below, packaged plants are available although costs for these plants are not presented here.

Capital costs, shown in Figure 2-22, are expressed as the total capital investment for a new granulated activated carbon treatment system as a function of design capacity. Net annual operating expenses are given in Figure 2-23, and unit annualized costs are shown in Figure 2-24. The cost curve of Figure 2-22 was plotted from estimates for three plant sizes: 2 mgd, 20 mgd, and 110 mgd design capacity. Pressure steel contactors were assumed for the 2 mgd and 20 mgd plants, and gravity steel contactors for the 110 mgd plant. Because on-site regeneration may not be economically justified for small plants, off-site

Table 2-3. Design Parameters for Typical Granular Activated Carbon Systems for Drinking Water Treatment (2,4,5)

Design flow rate, mgd	3	30	150
Operating flow rate, mgd	2	20	110
Contactor type	Steel pressure vessel	Steel pressure vessel	Steel gravity vessel
Empty bed contact time, min.	20	20	20
Hydraulic loading, gpm/ft^2	5	5	5
Backwash pumping rate, gpm/ft^2	10	10	10
Regeneration frequency, months	2	2	2
Carbon losses, %	7	7	7
Regeneration method	Off-site 20 miles	On-site infra-red	On-site multi-hearth

Figure 2-22. Granular activated carbon system for drinking water treatment - Total capital investment (March, 1980 dollars).

Costs based on regeneration frequency of 2 months. Costs below 2 mgd are by extrapolation.

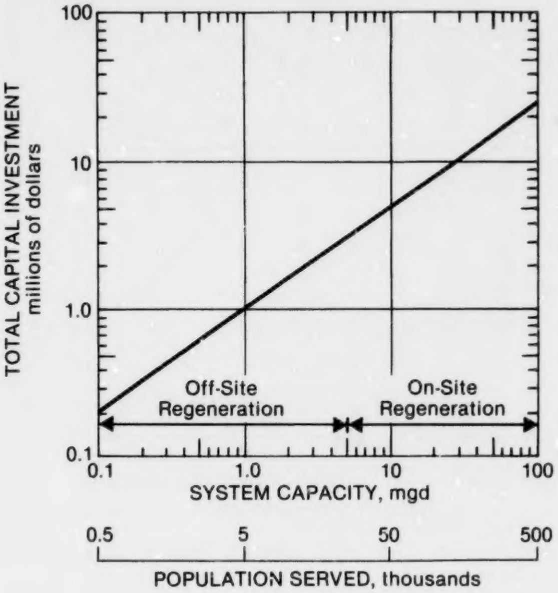


Figure 2-23. Granular activated carbon system for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

Costs based on regeneration frequency of 2 months. Costs below 2 mgd are by extrapolation.

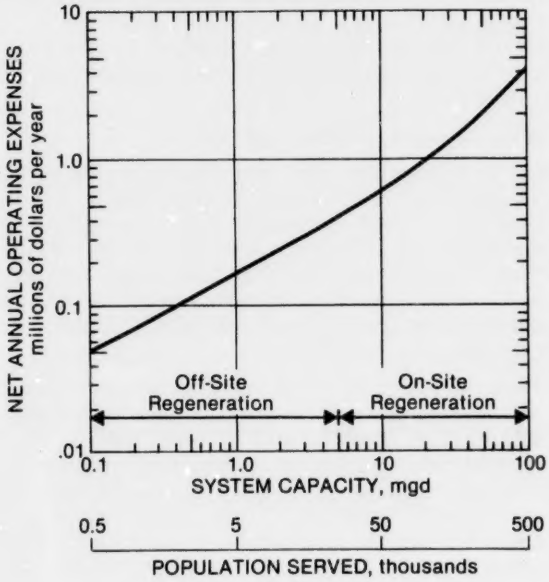
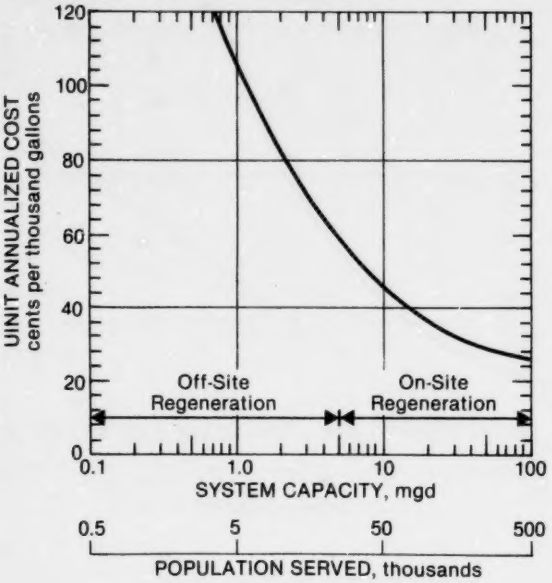


Figure 2-24. Granular activated carbon system for drinking water treatment - Unit annualized cost (March, 1980 dollars).

Costs based on regeneration frequency of 2 months. Costs below 2 mgd are by extrapolation.



regeneration is the basis for costs of smaller plants as indicated. Five percent of the cost of an off-site regional regeneration facility is apportioned to the 2 mgd plant. On-site multiple hearth furnace regeneration is assumed for the 20 and 110 mgd plants. The size ranges for the two regeneration options are indicated on the cost curves. Actually there is not an abrupt change at a particular size, but a range over which the relative attractiveness of off-site versus on-site regeneration must be compared on a case by case basis.

Major Variables Affecting Costs

Other than plant flow capacity, the major factors which affect overall capital costs for a granular activated carbon treatment system are: kind and concentration of organics in the influent water which determines required contact time in the contactors and hence contactor volume, carbon loading (lb contaminants adsorbed per lb of carbon), and regeneration frequency.

There is a cost trade-off between contactor volume and regeneration frequency. If smaller contactor volumes are used to reduce investment costs for the carbon contactor system, more frequent regeneration and the associated higher costs for regeneration equipment are incurred. Likewise, regeneration frequency and associated costs can be reduced by using larger contactors. An investigation of the effect of regeneration frequency on costs suggests that a regeneration interval of about 2 to 3 months is reasonable for granulated activated carbon systems in drinking water treatment (9).

Net annual operating expenses are strongly influenced by regeneration frequency, carbon losses, and fuel costs (10).

Carbon losses occur during handling of the carbon in charging and discharging equipment and during regeneration. Handling losses occur due to spillage and gradual attrition. Regeneration losses result when some of the carbon is burned along with the adsorbed organics during regeneration. Carbon losses typically range from 5 to 10 percent; in this cost summary they were assumed to be 7 percent.

2.4 Aeration

Description

Aeration is a process for removal of volatile organic materials from drinking water. Flowing streams of air and water are contacted with each other so that volatile organic materials are evaporated into the air stream and removed from the water. Aeration can be carried out in towers or aeration basins to provide the necessary contact between air and water. An aeration basin is typically constructed of concrete. An aeration tower is a rectangular structure similar to the water cooling towers used with large air conditioning systems. The two process options are illustrated conceptually in Figure 2-25.

For basin aeration, the water enters one or more open concrete contact basins. Compressed air is fed to air diffuser pipes set in the bottom of the basins. Air bubbles strip organic compounds from the water as they rise to the surface. A basin is designed to allow sufficient detention time for the air to reduce the concentration of organic compounds in the water.

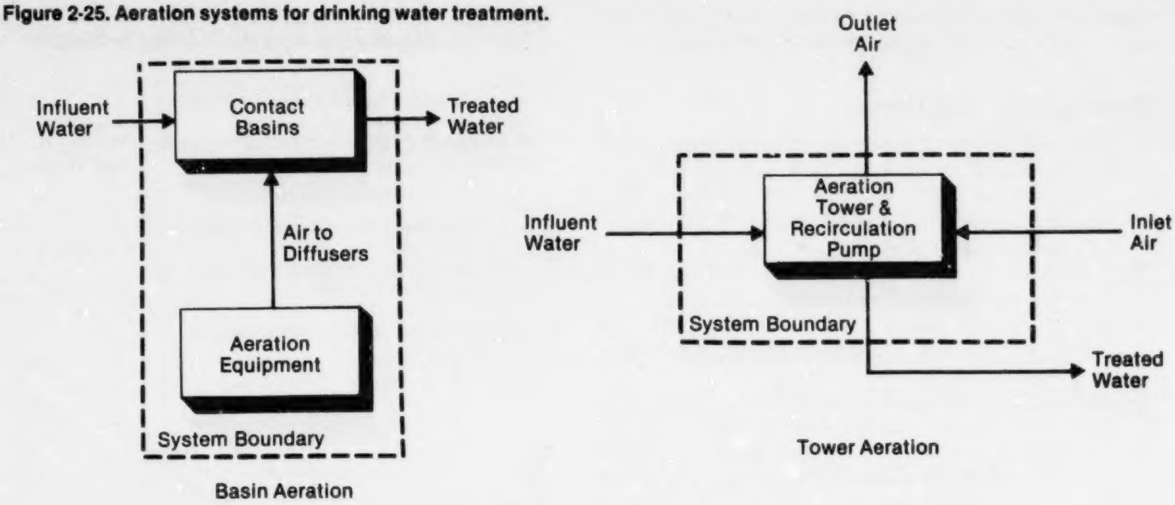
Similar to conventional cooling towers, aeration towers might consist of a fiberglass-covered metal framework containing a plastic packing medium. As water introduced near the top of the tower flows downward through the packing, it contacts air flowing upward. An induced draft fan in the tower stack draws in air at the bottom of the tower. Organic materials stripped from the water leave with the exit air stream. Treated water collects in a concrete basin beneath the tower; from there it is pumped to storage.

Design Basis and Costs (2,4,8)

The most significant design parameter for both basins and towers is the air-to-water ratio. Table 2-4 shows the major tower and basin design parameters used for the cost data presented here.

Costs for both basin and tower systems were derived for a conceptual design based on limited laboratory data (8). Test results for chloroform and several other

Figure 2-25. Aeration systems for drinking water treatment.



Basin Aeration

Tower Aeration

trihalomethanes were used to determine the sizes required for aeration basins and towers based on comparable performance in removing chloroform. Since the conceptual designs and cost estimates are based on limited laboratory data rather than pilot plant data, the cost estimates are very preliminary.

Table 2-4. Major Design Parameters for Aeration Basins and Towers for Trihalomethane[a] Removal in Drinking Water Treatment (4,9)

	Basins	
Performance	65% Removal	90% Removal
Air-to-water ratio	10:1	20:1
Number basins in series	1	2
Air loading, scfm/ft^2 of basin area	5	5
Operating temperature, °F	70-80	70-80
Basin depth, ft	12	12
Basin volume, ft^3/mgd	2,200	4,400
Influent chloroform concentration, mg/l	10-800	10-800

	Towers	
Performance	65% Removal	90% Removal
Air-to-water ratio	10:1	100:1
Typical water loading, mgd/ft^2 of tower cross-section	0.059	0.059
Design superficial air velocity, 1 ft/sec based on empty tower cross-section	1.0	1.0
Maximum superficial air velocity, 2.5 ft/sec based on empty tower	2.5	2.5
Tower water pumps (one operating, one spare) total dynamic head, ft	30	30
Operating temperature, °F	70-80	70-80
Tower volume, ft^3/mgd	340	3,400
Tower height, ft	22	22
Influent chloroform concentration, mg/l	10-800	10-800

[a] Although organic materials other than trihalomethanes can also be removed by aeration, the most data was available on THM removal. Also, at the time this report was written, a major interest was in THM removal. Thus, THM removal was the design basis for aerations.

The data showed that for 65 percent and 90 percent removal of chloroform in aeration towers the corresponding air-to-water ratios were 10 to 1 and 100 to 1, respectively. The data did not show a difference between basins and towers at the 10 to 1 ratio, but it appeared that the basins might achieve the higher removal at a ratio of 20 to 1. These design criteria were used as the basis for the costs in this report using cost data for basins and towers from References 4 and 5.

As a technology for the removal of organic compounds in general, the conditions observed in tests with trihalomethanes might be typical of conditions necessary for the removal of other organic materials so that these costs apply to basins and towers for the removal of other organic materials.

Total capital investment costs are shown in Figure 2-26, net annual operating expenses in Figure 2-27, and unit annualized cost in Figure 2-28. The two sets of curves reflect the two contact options of either basins or towers. The two curves in each set correspond to different air-to-water ratios (and hence removal efficiency).

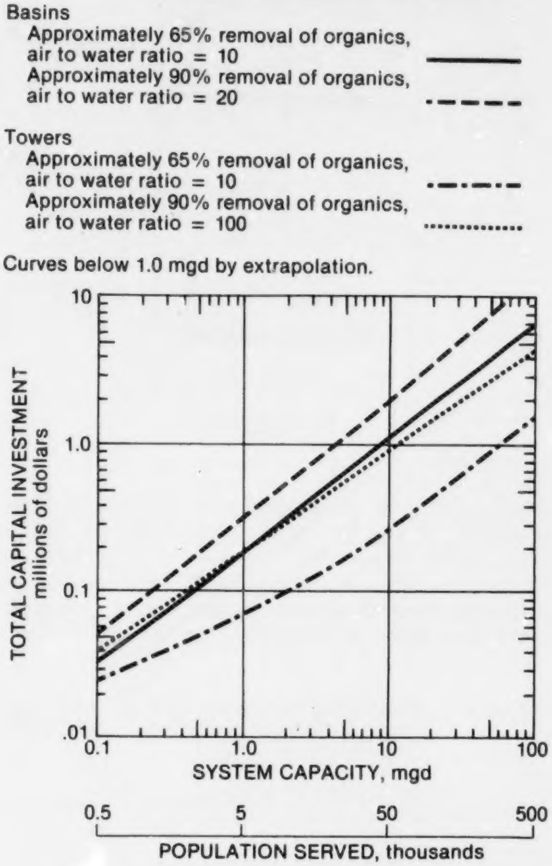
Major Variable Affecting Costs

Other than type of system (basin or tower), the most significant variable affecting costs for aeration is the air-to-water ratio required to achieve a specified level of performance. Air-to-water ratios that might be required to achieve comparable levels of performance in basins compared to towers can be expected to vary with concentrations and kinds of organic materials in different water supplies. The required ratio to achieve results is also sensitive to temperature with higher temperatures improving removal efficiency and lowering the required air-to-water ratio. Other variables include details of equipment design which could differ from those described in this report.

References - Section 2

1. Midwest Research Institute. Small System Water Treatment Symposium, EPA-570/9-79-021. Prepared for U.S. Environmental Protection Agency, Office of Drinking Water, Washington, DC, September 1979.
2. Gumerman, R.C., et al. (Culp/Wesner/Culp) Estimating Water Treatment Costs, Vol. 1. Summary, EPA-600/2-79-162a (PB80-139819*). U.S. Environmental Protection Agency, Cincinnati, OH, August 1979.
3. Clark, R.M. and R.G. Stevie. Meeting the Drinking Water Standards in Safe Drinking Water: Current and Future Problems. Proceedings of a National Conference, Resources of the Future. Research Paper 12. Washington, DC, Clifford S. Russell, (ed.) 1978.
4. Gumerman, R.C., et al. (Culp/Wesner/Culp) Estimating Water Treatment Costs, Vol. 2. Cost Curves Applicable to 1-200 mgd Treatment Plants, EPA-600/2-79-162b (PB80-139827*). U.S. Environmental Protection Agency, Cincinnati, OH, August 1979.
5. Hansen, S.P., et al. (Culp/Wesner/Culp) Estimating Water Treatment Costs, Vol. 3. Cost Curves Applicable to 2,500 gpd to 1 mgd Treatment Plants, EPA-600/2-79-162c (PB80-148455*). U.S. Environmental Protection Agency, Cincinnati, OH, August 1979.
6. U.S. Environmental Protection Agency, Cincinnati, OH. Process Design Manual for Phosphorus Removal. EPA-625/1-76-001a (PB259150*), April 1976.
7. U.S. Environmental Protection Agency, Cincinnati, OH. Process Design Manual for Suspended Solids Removal. EPA-625/1-75-003a (PB259147*), January 1975.
8. Symons, J.M., et al. Handbook of Treatment Techniques for the Control of Trihalomethanes in Drinking Water. U.S. Environmental Protection Agency, Office of Research and Development, Municipal Environmental Research Laboratory, Cincinnati, OH, May 1980. (In preparation)
9. Clark, R.M. and P. Dorsey. Influence of Operating Variables on the Cost of Granular Activated Carbon Adsorption Treatment, EPA-600/d-78-001. U.S. Environmental Protection Agency, Cincinnati, OH, 1979.
10. Clark, R.M., et al. The Cost of Removing Chloroform and Other Trihalomethanes from Drinking Water Supplies, EPA-600/1-77-008 (PB264283*). U.S. Environmental Protection Agency, Cincinnati, OH, March 1977.

Figure 2-26. Aeration for drinking water treatment - Total capital investment (March, 1980 dollars).



Curves below 1.0 mgd by extrapolation.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

Figure 2-27. Aeration for drinking water treatment - Net annual operating expenses (March, 1980 dollars).

Basins
  Approximately 65% removal of organics, air to water ratio = 10 ———
  Approximately 90% removal of organics, air to water ratio = 20 – – – –

Towers
  Approximately 65% removal of organics, air to water ratio = 10 -·—·—·
  Approximately 90% removal of organics, air to water ratio = 100 ··········

Curves below 1.0 mgd by extrapolation.

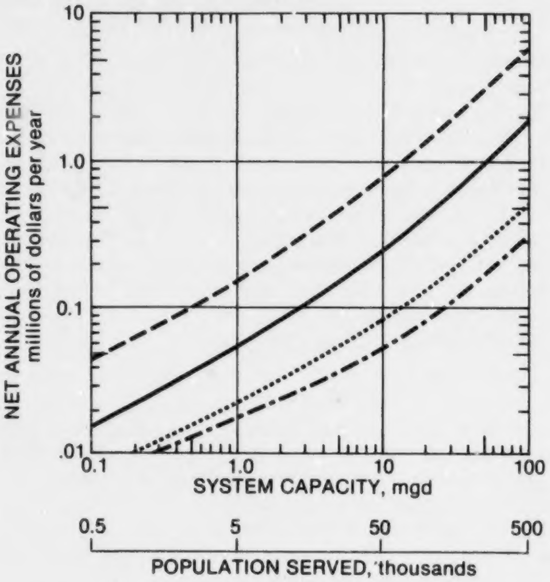
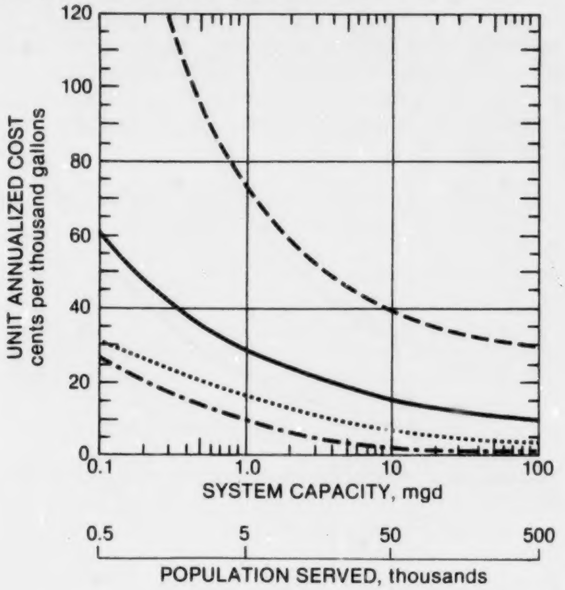


Figure 2-28. Aeration for drinking water treatment - Unit annualized cost (March, 1980 dollars).

Basins
  Approximately 65% removal of organics, air to water ratio = 10 ———
  Approximately 90% removal of organics, air to water ratio = 20 – – – –

Towers
  Approximately 65% removal of organics, air to water ratio = 10 -·—·—·
  Approximately 90% removal of organics, air to water ratio = 100 ··········

Curves below 1.0 mgd by extrapolation.



Section 3
Wastewater Treatment

Wastewater from domestic, municipal, and industrial sources must be treated to remove pollutants that are harmful to human health and the environment. Of the more than 300 billion gallons of water drawn for use in the United States each day, approximately 90 percent is used by industry (1). Although some of this quantity is lost through evaporation or incorporated into products, a substantial portion is discharged as wastewater.

The four major categories of wastewater sources are:

- Steam electric power generation.
- Agriculture.
- Manufacturing and minerals production.
- Domestic, commercial, and public sources.

These sources comprise a variety of wastewater stream characteristics requiring different types of treatment prior to discharge. Wastewater is treated in more than 2,500 municipal treatment plants in the U.S. (2), as well as in numerous industrial facilities.

Historically, the body of legislation dealing with wastewater discharges has increased gradually. But increased national attention to this issue occurred in the last decade. Although the federal government had been active in funding municipal wastewater treatment facilities since 1957 (2), the legislative initiatives of the 1970's provided additional impetus for wide-scale cleanup of the nation's waterways. The major recent legislation was the Clean Water Acts of 1972 and 1977.

General health, aesthetic, and recreational reasons were the early sources of motivation for water cleanup. Prevention of long-term uncertain deleterious effects on health and the environment is now a growing consideration. Increasing demands on the nation's water resources will likely increase water reuse in many areas of the country. This would increase the use of wastewater treatment technologies.

Total capital expenditures for water pollution control were reported as $10.9 billion in 1977. Annualized costs were reported as $8.9 billion. Total investment spending for water pollution control between 1977 and 1986 has been estimated at $50.7 billion in constant 1977 dollars. Total annualized cost expenditures have been estimated as $121.8 billion for the same period (3).

Several categories of water pollutants are of interest. The water pollutants controlled by the technologies in this section are organic substances, suspended solids, phosphorus containing compounds (both suspended and dissolved), and ammonia.

Organic waste concentrations are commonly expressed as 5-day biochemical oxygen demand (BOD_5), chemical oxygen demand (COD), or total organic carbon (TOC) in milligrams per liter (mg/l). Historically, BOD_5 has been used to express the biodegradable waste concentration in municipal as well as industrial wastewater. COD provides a measure of the presence of refractory organic materials not amenable to biological treatment. TOC measures total organic materials. These measures of organic waste concentrations include both dissolved materials and suspended solids.

Suspended solids (SS) include both organic and inorganic, biologically inert materials such as fine particles of silt. Organic suspended solids contribute to a portion of the total BOD_5 of the wastewater. Phosphorus (P) is present as dissolved phosphorus compounds as well as in some of the suspended solids. Ammonia (NH_3) is present both as dissolved ammonia gas and in soluble compounds. Ammonia can form from the degradation of nitrogeneous compounds in the waste. Concentrations of all these waste materials are usually expressed as milligrams per liter (mg/l). Additional pollutants commonly removed from wastewater are bacteria, viruses, and soluble minerals which interfere with subsequent intended uses of the treated wastewater.

Treatment technologies for removing the pollutants discussed above with corresponding percentage removal capabilities are listed in Table 3-1. The first

Table 3-1. Typical Pollutant Removal Efficiency of Wastewater Treatment Technology (4,5)

	Pollutant removal efficiency, percent				
Technology	BOD_5	COD	SS	P	NH_3
Conventional secondary and advanced wastewater treatment	80-95	50-70	80-90	25-45	10-20
Stabilization ponds and aerated lagoons	60-90	70-90	70-90	25-30	25-95
Land treatment	95-99	—	95-99	≤90	>25
Phosphorus removal by chemical addition	—	—	—	90-95	—
Nitrification	—	—	—	—	≥98
Granular media filtration	—	—	>95	—	—

three technologies listed are biological treatment systems comprising several process steps which remove a portion of all of the pollutants listed. The last three systems are individual process steps which specifically remove phosphorus, ammonia, and suspended solids, respectively, although some reduction in the other pollutants also occurs.

As of 1977, 737 wastewater treatment projects funded under the Municipal Wastewater Treatment Plants Grants Program and catalogued as part of a cost review (6) were distributed as follows:

Activated sludge*, %	47.1
Stabilization ponds, %	11.5
Aerated lagoons, %	3.9
Other, %	37.5

Some of the systems in the "other" category included trickling filter plants and rotating biological contactors. The remaining kinds of systems were not identified.

For municipal wastewater treatment technologies in this report, the typical process design and corresponding costs are based on the following influent wastewater characteristics:

BOD_5, mg/l	210
Suspended solids, mg/l	230
Total phosphorus (as P), mg/l	11
Ammonia, mg/l	19
pH	7.0

There are wide variations in industrial wastewater characteristics. The characteristics listed above for municipal wastewater are also typical for some cases of industrial wastes at the low end of the BOD_5 concentration range. In these cases, industrial treatment costs will be similar to those for municipal wastewater. However, since typical industrial BOD_5 concentrations can be much greater than 210 mg/l, the process designs in this section also consider an influent wastewater with a BOD_5 level of 1000 mg/l. It should be recognized that this is only a special case for industrial wastewater treatment. This particular BOD level was selected to illustrate the effect of this parameter on cost.

3.1 Conventional Secondary and Advanced Wastewater Treatment

Description

Wastewater treatment processes that achieve effluent levels of 30 mg/l or less of 5-day biochemical oxygen demand, BOD_5, and 30 mg/l or less of suspended solids are referred to as conventional secondary treatment (6). Those systems which achieve effluent levels of 10 mg/l or less of BOD_5, and 10 mg/l or less of suspended solids are referred to as advanced wastewater treatment (6). Both types of treatment systems can use a number of combinations of unit processes to achieve these effluent levels. Advanced wastewater treatment plants use the same process operations as conventional secondary treatment plants with additional processing steps to achieve greater removal of pollutants. Individual treatment plants of either kind can differ in details of component equipment configurations and specifications because of differences in influent water characteristics, treatment objectives, and other site-specific considerations.

The typical conventional secondary treatment system considered in this report contains the following major process modules:

- Preliminary treatment.
- Influent pumping.
- Primary clarification.
- Activated sludge secondary treatment.
- Secondary clarification.
- Effluent disinfection by chlorination
- Sludge treatment.

A typical advanced wastewater treatment system contains, in addition to the above, the following process modules:

- Primary chemical addition (prior to primary clarification).
- Secondary chemical addition (prior to secondary clarification).
- Granular media filtration of secondary clarifier effluent.

An additional process module, granular activated carbon treatment, could be used after granular media filtration, but is not considered here.

Configurations of typical systems for conventional secondary treatment and advanced wastewater treatment are shown conceptually in Figure 3-1 and Figure 3-2, respectively.

Influent enters a preliminary treatment module where debris and large suspended solids such as grit are removed. Sometimes, the flow in preliminary treatment is equalized in a large holding basin to dampen the effect of fluctuations in influent flow rates and waste loadings on downstream process modules. Flow equalization enhances the downstream removal of contaminants by providing a more uniform waste stream.

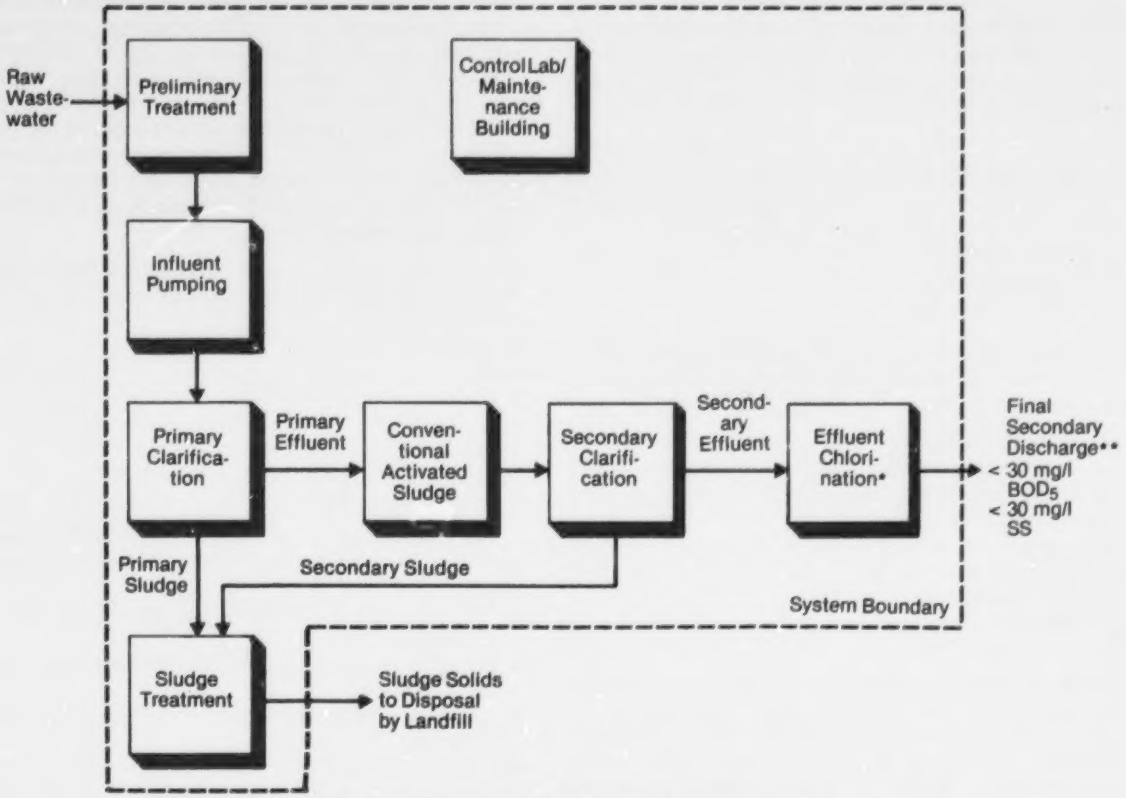
Effluent from preliminary treatment flows to the primary clarifiers. The clarifiers provide a relatively long detention time so that a large portion of the suspended solids can settle out. Chemical coagulants and coagulant aids can be used to enhance the removal of solids. Conventional systems sometimes use chemicals; advanced treatment systems nearly always use chemicals in this step.

Clarifiers can be either rectangular or circular, and fabricated of either concrete or steel. Sludge (settled

---

*The common biological treatment step in conventional secondary and advanced wastewater treatment.

Figure 3-1. Conventional secondary treatment system for wastewater.


Raw Waste-water
Preliminary Treatment
Control Lab/ Maintenance Building
Influent Pumping
Primary Clarification
Primary Effluent
Conventional Activated Sludge
Secondary Clarification
Secondary Effluent
Effluent Chlorination*
Final Secondary Discharge**
< 30 mg/l BOD_5
< 30 mg/l SS
Primary Sludge
Secondary Sludge
System Boundary
Sludge Treatment
Sludge Solids to Disposal by Landfill

*Effluent chlorination is not commonly used for industrial wastes.

**These effluent concentrations do not necessarily apply to industrial wastewater, but are characteristic of effluent discharges from municipal wastewater treatment.

suspended solids) is removed from the bottom of the clarifier vessel and is pumped to sludge treatment. Clarifier effluent flows to the activated sludge aeration tanks for further treatment.

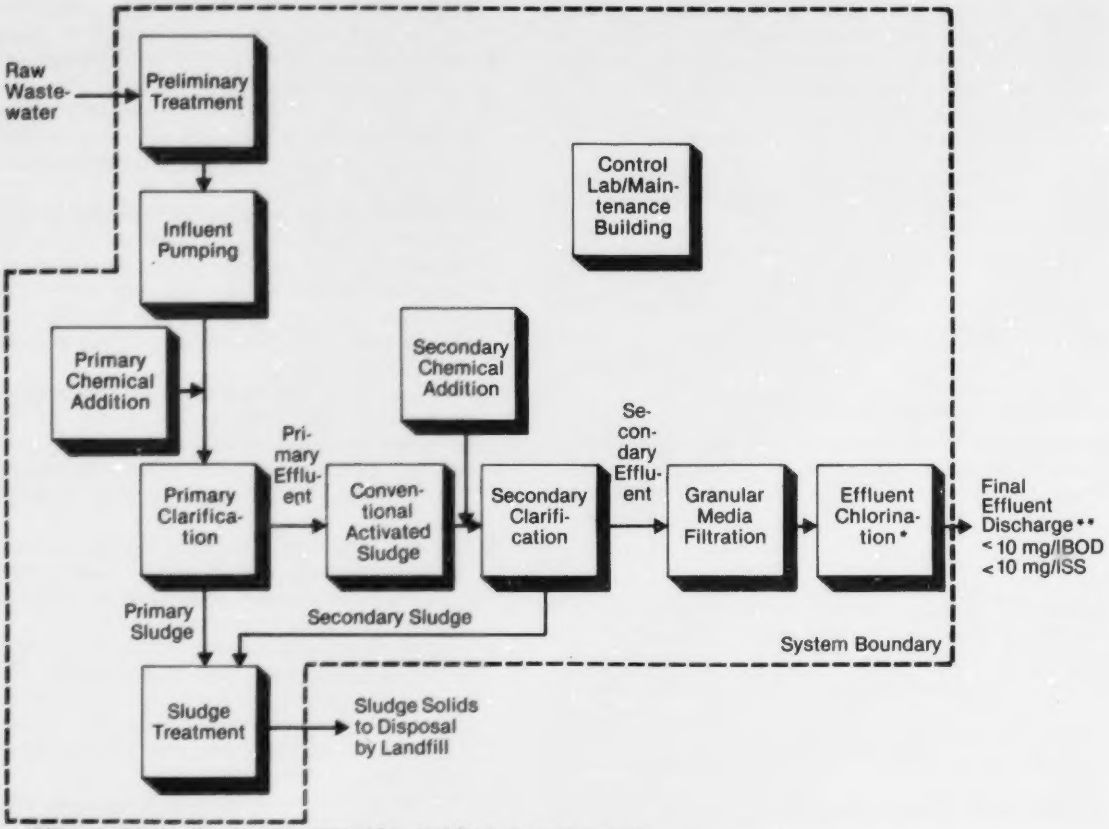
Conventional activated sludge treatment is a continuous-flow biological process. A suspension of aerobic microorganisms is mixed into the wastewater; the mixture of microorganisms and wastewater, called mixed liquor, is agitated by air bubbles rising from diffuser pipes in the bottom of the aeration vessel or by mechanical surface aerators. The microoganisms oxidize soluble and colloidal organic compounds to carbon dioxide and water. The mixture flows from the aeration vessel to secondary clarifiers for separation of solids. These clarifiers are similar to the primary clarifiers discussed above.

Secondary clarifiers remove some of the suspended solids from the activated sludge aeration vessel effluent. A portion of the solids settled out in the secondary clarifiers is returned to the aeration tank inlet as recycle sludge to seed biological activity in the incoming wastewater. Excess sludge resulting from microorganism growth is routed to the sludge treatment processes for disposal.

In the clarifiers chemical addition can be used to enhance settling. In conventional secondary treatment, the clarified secondary effluent may be disinfected prior to discharge. In advanced wastewater treatment, the secondary effluent passes through granular media filters which further reduce suspended solids and BOD_5 to the required advanced wastewater treatment levels ($BOD_5 \leq 10$ mg/l, SS ≤ 10 mg/l for municipal wastewater).

For some high strength industrial wastes, some of the secondary clarifier effluent is recycled to the activated sludge aeration vessels in order to dilute high levels of BOD_5. Lower BOD_5 levels in the aeration vessel may be necessary to ensure the

Figure 3-2. Advanced wastewater treatment system.


Raw Wastewater
Preliminary Treatment
Influent Pumping
Control Lab/Maintenance Building
Primary Chemical Addition
Secondary Chemical Addition
Primary Effluent
Secondary Effluent
Primary Clarification
Conventional Activated Sludge
Secondary Clarification
Granular Media Filtration
Effluent Chlorination *
Final Effluent Discharge ** < 10 mg/lBOD < 10 mg/lSS
Primary Sludge
Secondary Sludge
System Boundary
Sludge Treatment
Sludge Solids to Disposal by Landfill

*Effluent chlorination is not commonly used for industrial wastes.
**These effluent concentrations do not necessarily apply to industrial wastewater, but are characteristic of effluent discharges from municipal wastewater treatment.

required removal efficiency. Another approach with high strength wastes is to provide a longer detention time in the aeration vessel than for low strength wastes.

Sludge treatment is used to reduce the volume of sludge from both primary and secondary clarifiers and to render the sludge more acceptable for final disposal. A number of sludge treatment options may be used. One common method is thickening, digestion, dewatering, and final disposal by landfill.

Secondary clarifier sludge, which contains about 95 percent water, is commonly concentrated in a gravity thickener. From this process the sludge is transferred to a digester which chemically and physically alters the sludge solids to facilitate ultimate disposal.

Sludge digestion can be either aerobic or anaerobic. Anaerobic digestion, which is most commonly employed, converts sludge into methane, carbon dioxide, and a residual organic material. The digestion takes place in the first of two tanks in series. The second tank provides for settling of solids and separation of supernatant liquid which is routed to a previous process step. Combustible gas is collected from both stages and used as heater fuel in the treatment plant. Sludge is dewatered to increase the solids content prior to final disposal.

Dewatering can be accomplished by sand-bed drying, vacuum filtration, or centrifugation, depending on the physical properties of the sludge. Landfill, incineration, land spreading, and other methods are used for final dewatered sludge solids disposal.

Conventional secondary or advanced wastewater treatment using the activated sludge process can be applied to both domestic wastewater and biodegradable industrial wastewater. It is not uncommon for municipal and industrial wastewaters to be combined for treatment. In these cases the industrial waste cannot contain toxic materials that would render the biological treatment process inoperative or refractory materials that would result in effluent standards being exceeded. Also, the industrial waste might require special provisions for oil and grease separation as part of preliminary treatment.

Advanced wastewater treatment achieves higher quality effluent than can be achieved by conventional secondary treatment. If the non-biodegradable organic portion of the waste is large enough to cause problems in receiving water bodies, granular activated carbon treatment might be required to reduce effluent organic concentrations.

Design Basis and Costs (4,6,7,8)

As shown in Table 3-2, one typical municipal and two typical industrial wastewater compositions were selected as the design basis for conventional secondary and advanced wastewater treatment systems.

Characteristics of industrial wastes vary widely, depending on both the particular industry and the individual operating facility. High BOD_5 affects the design of the activated sludge units, secondary clarifiers, and sludge treatment. At high BOD_5 levels, the volume of excess sludge from microbial growth during the activated sludge process is much greater than any reasonable level of inert suspended solids likely in the raw waste influent. For a given plant capacity, activated sludge units, secondary clarifiers, and sludge treatment systems must all be larger for high BOD_5 levels than for low BOD_5 levels. This is because the volume of sludge generated by microbial growth increases with BOD_5 level if the same food (BOD_5) to microorganism ratio (F/M) is maintained. The volume of sludge generated by microbial growth was based on a reported value from the literature (9). This value, the influent concentrations of Table 3-2, and typical removal efficiencies for unit processes were used to develop the system material balance for sizing each unit process.

Key design parameters for the conventional secondary treatment process are given in Table 3-3 and for the advanced wastewater treatment process in Table 3-4. Some design features for the treatment systems are outlined below:

Table 3-2. Typical Influent Wastewater Composition

	Municipal or medium strength industrial	High strength industrial
BOD_5, mg/l	210	1000
SS, mg/l	230	230
Totol phosphorus (as P), mg/l	11	—
Ammonia nitrogen, mg/l	19	—
pH	7.0	7.0

- The preliminary treatment module contains a bar screen and grit chamber. The grit chamber is a horizontal flow type with mechanical grit handling.
- Influent pumping capacity is provided for twice the overall plant design flow.
- Circular primary clarifiers have been specified. Primary sludge pumps are included to transport the settled solids to the sludge treatment portion of the overall system.
- Activated sludge aeration vessels are rectangular concrete basins sized appropriately for the required detention time. Diffused aeration is used. These are followed by circular secondary clarifiers provided with sludge pumps to transfer solids to sludge treatment. Secondary sludge is combined with primary sludge.
- For the advanced wastewater treatment plant, granular media filter units consist of multiple concrete or steel vessels containing a sand bed overlain with a bed of anthracite. Total bed depth is from 2 to 5 ft. These units include backwash systems.
- Effluent from the secondary clarifiers in conventional treatment and granular media filters in advanced wastewater treatment is disinfected by chlorination prior to final discharge.
- Sludge treatment includes thickening, digestion, and dewatering, with final disposal by landfill. The sludge thickeners are circular tanks similar to the clarifiers and include discharge pumps. Two-stage anaerobic digestion is assumed. Sludge dewatering is by vacuum filter.
- All piping and miscellaneous pumps, electrical equipment, instrumentation, required service auxiliaries, and buildings are included.

Two sets of cost curves for total capital investment, plotted against plant capacity in millions of gallons per day, are shown in Figure 3-3 corresponding to conventional secondary treatment and advanced wastewater treatment. Two curves in each set correspond to different influent BOD_5 levels for municipal or medium strength industrial wastewater and high strength industrial wastewater. In all cases, final disposal of sludge solids is by landfill. Net annual operating expenses that correspond to the capital investment curves are shown in Figure 3-4. Unit annualized cost is given in Figure 3-5.

Major Variables Affecting Costs

Among the variables that could significantly affect the costs of conventional secondary and advanced wastewater treatment are variations in individual wastewater characteristics and the equipment sizing changes that would occur as a result. The complexity of these wastewater treatment systems makes a quantitative analysis of such effects beyond the scope of this discussion.

Table 3-3. Design Parameters for Conventional Secondary Treatment System (4,6,7)

Parameter	Units	Range	Design value[a]
Primary clarifiers			
Surface loading	gpd/ft^2	600-1,200	800
Detention time	hr	1.5-3.0	N/A
Activated sludge aeration vessels			
Volumetric loading	lb BOD_5/day/1000 ft^3	25-30	32
Detention time	hr	4-38	6-29
MLVSS[b]	mg/l	1,500-3,000	2,100
F/M ratio[c]	lb BOD_5/day/lb MLVSS	0.25-0.5	0.25
Air requirement	ft^3/lb BOD_5 removed	700-1,500	700
Secondary clarifiers			
Surface loading	gpd/ft^3	400-800	600
Gravity thickener			
Solids loading	lb/ft^2/day	4-8	6
Sludge digester			
Solids loading	lb VSS/ft^3/d[d]	0.04-0.40	0.16
Operating temperature	°F	85-110	85-110
Sludge dewatering (by vacuum filter)			
Sludge solids concentration	lb solids/10^{6}gal	-	900
Dry solids loading	lb solids/hr/ft^2	3.5-15	5
Operating schedule			
1 mgd plant	hr/day	-	6
10 mgd plant	hr/day	-	12
100 mgd plant	hr/day	-	10
Chemical treatment dosage			
$FeCl_3$	lb/10^{6}gal	-	35
CaO	lb/10^{6}gal	-	90

[a] N/A = Not available in cost reference used.

[b] MLVSS = Mixed liquor volatile suspended solids in the aeration vessel, a measure of microorganism population.

[c] F/M = Food to microorganism ratio, a measure of organic waste concentration to microorganism population.

[d] VSS = Volatile suspended solids, in the digester; a measure of digestible solids which can be converted to CO_2 and H_2O.

Table 3-4. Design Parameters for Advanced Wastewater Treatment System (4,6,7)

Parameter	Units	Range	Design Value
Primary clarifiers			
Surface loading	gpd/ft^2	600-1,200	800
Detention time	hr	1.5-3.0	NA[a]
Activated sludge aeration vessels			
Volumetric loading	lb BOD_5/day/1,000 ft^3	25-50	32
Detention time	hr	4-38	6-29
MLVSS[b]	mg/l	1,500-3,000	2,100
F/M ratio[c]	lb BOD_5/day/lb MLVSS	0.25-0.5	0.25
Air requirement	ft^3/lb BOD_5 removed	700-1,500	700
Secondary clarifiers			
Surface loading	gpd/ft^3	400-800	600
Gravity thickener			
Solids loading	lb/ft^2/day	4-8	6
Sludge digester			
Solids loading	lb VSS/ft^3/d[d]	0.04-0.40	0.16
Operating temperature	°F	85-110	85-110
Sludge dewatering (by vacuum filter)			
Sludge solids concentration	lb solids/10^{6}gal	-	900
Dry solids loading	lb solids/hr/ft^2	3.5-15	5
Operating schedule			
1 mgd plant	hr/day	-	6
10 mgd plant	hr/day	-	12
100 mgd plant	hr/day	-	10
Chemical treatment dosage			
$FeCl_3$	lb/10^{6}gal	-	35
CaO	lb/10^{6}gal	-	90

(Continued)

Table 3-4. Continued.

Parameter	Units	Range	Design Value
Primary and Secondary			
Chemical addition			
Alum dosage[e]	mg/l	100-500	100
Granular media filtration[f]			
Hydraulic loading	gpm/ft^2	2-8	4
Run length	hours	8-48	12
Backwash cycle time	min.	-	15
Backwash hydraulic loading	gpm/ft^2	15-25	15

[a] N/A = Not available in cost reference used.
[b] MLSS = Mixed liquor volatile suspended solids in the aeration vessel, a measure of microorganism population.
[c] F/M = Food to microorganism ratio, a measure of organic waste concentration to microorganism population.
[d] VSS = Volatile suspended solids; in the digester, a measure of digestible solids which can be converted to CO_2 and H_2O.
[e] Refer to Section 3.4 of this report.
[f] Refer to Section 3.6 of this report.

Figure 3-3. Conventional secondary and advanced wastewater treatment - Total capital investment (March, 1980 dollars).

AWT - Industrial Waste (BOD = 1000 mg/l) ———
AWT - Municipal Industrial Waste (BOD = 210 mg/l) — — —
CST - Industrial Waste (BOD = 1000 mg/l) —·—·—
CST - Municipal Industrial Waste (BOD = 210 mg/l) ··········

Costs based on sludge dewatering by vacuum filtration.

For conventional secondary plants of 1 mgd and below, sludge dewatering by drying beds would reduce costs by about 33%.

For conventional secondary plants of 10 mgd and above, incineration of sludge would increase costs approximately 16% for waste at 210 mg/l BOD and 7% for waste at 1000 mg/l BOD.

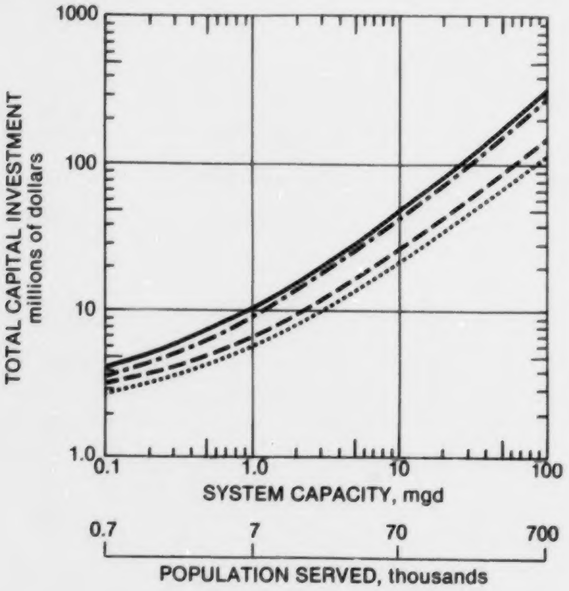


Figure 3-4. Conventional secondary and advanced wastewater treatment - Net annual operating expenses (March, 1980 dollars).

AWT - Industrial Waste (BOD = 1000 mg/l) ———
AWT - Municipal Industrial Waste (BOD = 210 mg/l) — — —
CST - Industrial Waste (BOD = 1000 mg/l) —·—·—
CST - Municipal Industrial Waste (BOD = 210 mg/l) ··········

Costs based on sludge dewatering by vacuum filtration.

For conventional secondary plants of 1 mgd and below, sludge dewatering by drying beds would reduce costs by about 23%.

For conventional secondary plants of 10 mgd and above, incineration of sludge would increase costs approximately 25% for waste at 210 mg/l BOD and 18% for waste at 1000 mg/l BOD.

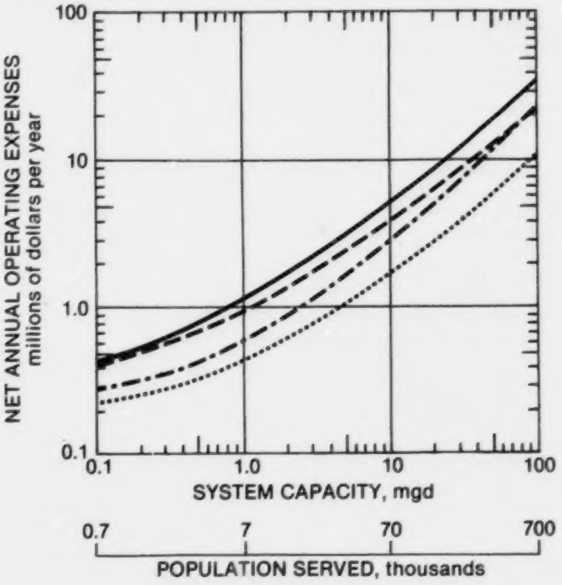


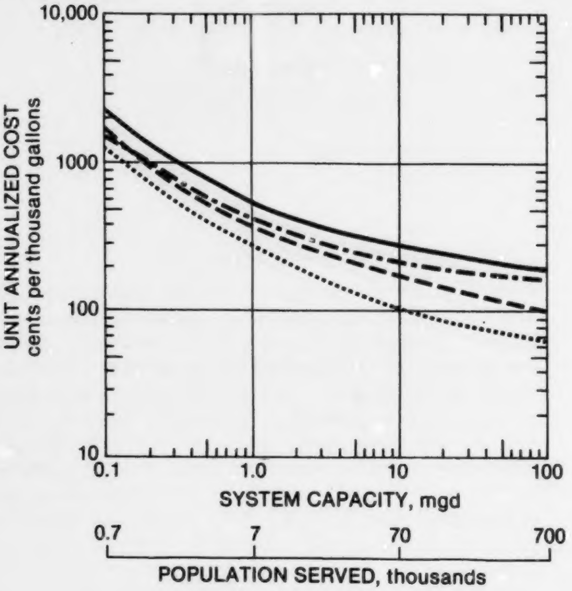
Figure 3-5. Conventional secondary and advanced wastewater treatment - Unit annualized cost (March, 1980 dollars).

AWT - Industrial Waste (BOD = 1000 mg/l) ———
AWT - Municipal Industrial Waste (BOD = 210 mg/l) — — —
CST - Industrial Waste (BOD = 1000 mg/l) –·–·–
CST - Municipal Industrial Waste (BOD = 210 mg/l) ·········

Costs based on sludge dewatering by vacuum filtration.

For conventional secondary plants of 1 mgd and below, sludge dewatering by drying beds would reduce costs by about 30%.

For conventional secondary plants of 10 mgd or above, incineration of sludge would increase costs approximately 20% for waste at 210 mg/l BOD and 11% for waste at 1000 mg/l BOD.



Any of a large number of significant factors can impact costs, depending on the value of each variable. Some of these variables are:

- Influent waste loadings.
- Solid settling characteristics.
- Specific chemical composition of the waste.
- Operating temperatures.
- Aeration methods.
- Sludge properties.
- Sludge treatment methods.
- Plant labor, energy and maintenance requirements.

In addition individual architectural features and plant layout can be a significant factor for these plants because of the large number of process steps. Individual plant administrative and operating practices can significantly affect operating expenses.

3.2 Stabilization Ponds and Aerated Lagoons

Description (4,6,7,10)

Ponds, or lagoons, are earthwork structures which can be below grade, at grade with earthwork dikes, or built by damming a natural terrain depression. The ponds can be unlined or lined with relatively impermeable clay, rubber, or plastic. They can be subdivided by earthwork partitions into several compartments or cells which provides for flexibility of operation. This flexibility makes it possible to optimize effluent quality. For example, each compartment can operate separately, or there can be circulation between compartments. Compartments may operate in parallel or in series.

Ponds treat wastewater by providing detention time for biological oxidation of BOD_5 and settling of suspended solids. The settled solids undergo anaerobic decomposition at the bottom of the pond. Detention time depends on individual wastewater characteristics, pond waste loading (lb BOD_5/acre/day), and operating temperature.

Ponds can be divided into two general classifications: an impounding, or absorption pond or a flowthrough pond (10). An impounding or absorption pond relies on percolation and evaporation to accommodate continued wastewater additions to the pond. Intermittent discharge can occur when peak flows exceed the pond's surge capacity.

Flowthrough ponds are of four basic types:

- Aerobic algae ponds.
- Facultative ponds (aerobic upper layer and anaerobic lower layer).
- Anaerobic ponds.
- Aerated ponds (aerated lagoons).

The first three types can be referred to as stabilization ponds; although, in this report, the term is reserved for facultative ponds. The above ponds differ in functional characteristics which are discussed below.

Aerobic ponds rely on algae growth to provide the oxygen necessary to satisfy the wastewater BOD_5 requirement. The depth of these ponds is, consequently, restricted to less than 5 ft to permit sunlight penetration. Some mixing is required to ensure good oxygen distribution. These shallow ponds rely on natural circulation. Provisions must also be made for separating the algae from the treated water prior to discharge. Sometimes this is accomplished by overflow design and sometimes a separate earthwork compartment or clarifier is provided for solids' settling prior to effluent discharge.

A facultative pond (stabilization pond) contains an upper water layer which behaves in the same manner as an aerobic pond. The bottom water layer and sludge on the pond bottom provide an anaerobic environment where organic materials decompose to produce methane and other gases. Depths for these ponds range from about 3 to 6 ft.

Anaerobic ponds are anaerobic throughout their volume. They are relatively deep to minimize odor generating surface area and to retain heat so that anaerobiosis can proceed.

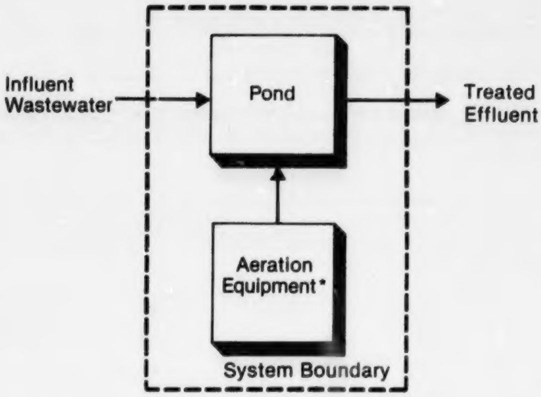
Aerated lagoons rely on mechanically promoted oxygen transfer to the wastewater. Diffused aeration or mechanical aeration systems can be used. These lagoons range in depth from 6 to 20 ft and can be subdivided into cells by earthwork partitions. Although upper layers of the pond are well aerated, anaerobic decomposition of solids does occur on the pond bottom. Surface mechanical aeration can be fix-mounted or floating, according to agitator design. The diffused air systems consist of perforated plastic pipes supported near the bottom of the cells. Regularly spaced sparger holes are drilled in the tops of the pipes.

A large fraction of the incoming solids and of the biological solids produced from waste conversion settle to the bottom of the lagoon cells. As the solids begin to accumulate, a portion will undergo anaerobic decomposition. Suspended solids removal is enhanced if the design includes several smaller aerated polishing cells following the last aerated cell. In some lagoon designs, when high-intensity aeration produces completely mixed (all aerobic) conditions, a final settling tank with solids recycle is required. Periodically solids must be removed from ponds and hauled to a landfill. Ponds are usually designed for years of service before cleanout is required.

Ponds can be used for both municipal and industrial wastewater where biological treatment is effective. Removal of BOD_5 ranges from about 60 percent to over 90 percent depending on wastewater characteristics and system design parameters. Ponds are commonly used where land is inexpensive and treatment costs and operational requirements are to be minimized.

A conceptual representation of pond technology is shown in Figure 3-6.

Figure 3-6. Stabilization pond or aerated lagoon system for wastewater treatment.



* Only for aerated lagoon.

Design Basis and Costs (2,4,11)

A typical pond system requires excavation and embankment construction, the seeding of earthwork slopes, embankment protection, hydraulic control works, aeration equipment (for aerated lagoons), and electrical equipment. Design of lagoons and ponds can be based on either detention time or BOD_5 loading per unit pond area. Within the range of detention times provided, an influent BOD_5 concentration of 210 mg/l results in the BOD_5 pond area loadings shown in Table 3-5. Costs are based on adjustments of published data for a BOD_5 loading of 1200 lb BOD_5/acre/day to determine the costs for different detention times. Stabilization ponds typically have longer detention times than do aerated lagoons because of lower rates of oxygen transfer and corresponding rates of waste treatment. Typical design parameters for pond systems are presented in Table 3-5 (4,7,10).

Table 3-5. Design Parameters for Pond Systems (4,7,10)

Parameter[a]	Aerated lagoons		Stabilization ponds	
	Range	Design value	Range	Design value
Influent BOD_5 loading, mg/l	—	210	—	210
Detention time, days	3-10	7	3-30	15
Depth, ft	6-20	15	0.6-10	3
Organic loading, lb BOD_5/acre/day	10-1200	1200	20-500	40
Power requirement, hp/10^6 gal capacity	30-40	36	—	

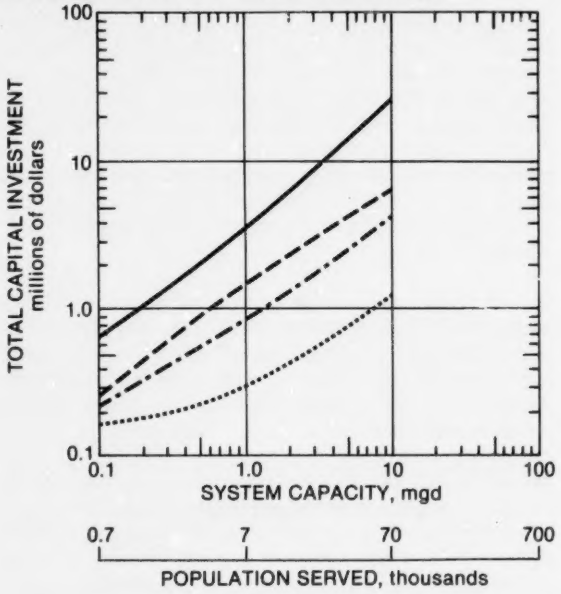
[a] Design basis for all cases includes oxygen requirement of 0.7 - 1.4 lb/lb BOD_5 removed.

Total capital investment is presented in Figure 3-7. Net annual operating expenses are presented in Figure 3-8 and unit annualized cost in Figure 3-9. The large variability that can occur in construction features and the corresponding effects on costs are so great for this technology that costs for site-specific cases should be expected to vary considerably from these cost curves.

Figure 3-7. Stabilization ponds and aerated lagoons for wastewater treatment - Total capital investment (March, 1980 dollars).

Stabilization Pond, Industrial Waste, BOD = 1000 mg/l ———
Stablization Pond, Municipal Waste, BOD = 210 mg/l – – – –
Pond BOD Loading = 40 lb BOD/acre/day

Aerated Lagoon, Industrial Waste, BOD = 1000 mg/l –·–·–
Aerated Lagoon, Municipal Waste, BOD = 210 mg/l ··············
Lagoon BOD Loading = 1200 lb BOD/acre/day



Major Variables Affecting Costs

As explained above, because a pond is primarily an earthwork construction, costs can be very site specific. Parker (11) listed the following major factors in determining investment cost:

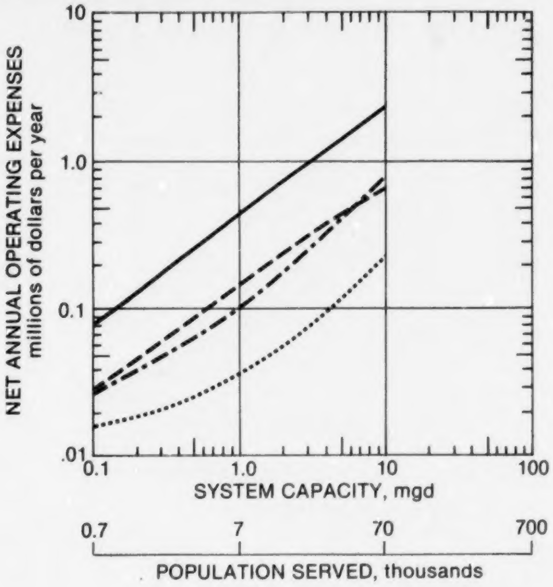
- Land availability and price.
- Pond surface area.
- Depth.
- Pond configuration.
- Terrain features.
- Dam or dike description.

Figure 3-8. Stabilization ponds and aerated lagoons for wastewater treatment - Net annual operating expenses (March, 1980 dollars).

Stabilization Pond, Industrial Waste, BOD = 1000 mg/l ———
Stabilization Pond, Municipal Waste, BOD = 210 mg/l – – – –
Pond BOD Loading = 40 lb BOD/acre/day

Aerated Lagoon, Industrial Waste, BOD = 1000 mg/l –·–·–
Aerated Lagoon, Municipal Waste, BOD = 210 mg/l ··············
Lagoon BOD Loading = 1200 lb BOD/acre/day

Net annual operating expenses excluding general expenses would be about 27% less for aerated lagoons and 70% less for stabilization ponds.



- Volume of earthwork involved.
- Type of earthwork.
- Pond lining requirements.
- Auxiliary construction.

The first cost item, the cost of the land itself, is highly sitespecific. Pond surface area is determined either from a specified pond depth and the average detention time of the wastewater impounded or from a permissible organic waste loading in the pond. These depend on individual wastewater characteristics.

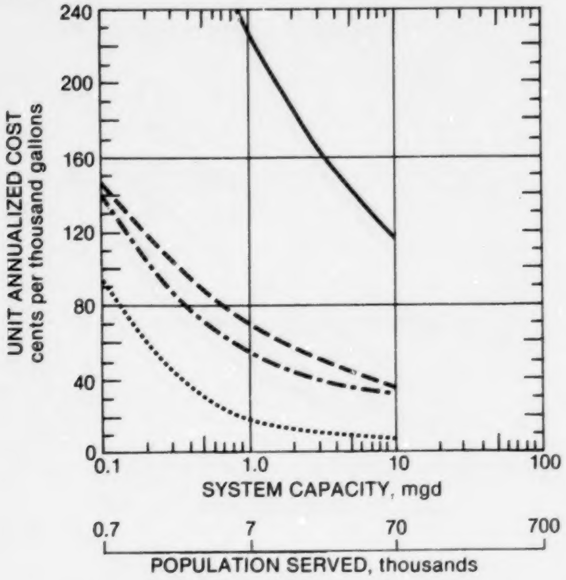
Depth significantly influences cost because of dike construction requirements. Dikes are wider at the bottom than at the top. As pond depth increases for a given surface area, the volume of earthwork in the dike increases by a greater amount than the increase in pond volume. Therefore, dike construction costs increase more rapidly as pond depth increases.

Figure 3-9. Stabilization ponds and aerated lagoons for wastewater treatment - Unit annualized cost (March, 1980 dollars).

Stabilization Pond, Industrial Waste, BOD = 1000 mg/l ———
Stabilization Pond, Municipal Waste, BOD = 210 mg/l - - - - -
Pond BOD Loading = 40 lb BOD/acre/day

Aerated Lagoon, Industrial Waste, BOD = 1000 mg/l —·—·—·
Aerated Lagoon, Municipal Waste, BOD = 210 mg/l ....
Lagoon BOD Loading = 1200 lb BOD/acre/day

Unit annualized cost based on net annual operating expenses without general expenses would be about 15% less for aerated lagoons and 46% less for stabilization ponds.



Pond configuration refers to shape and whether or not the pond is subdivided by earthwork partitions. Again there is a significant impact on the cost of dike construction. For a given surface area, dike perimeter increases in going from a square pond to a rectangular one. Because dike cross-sectional area is fixed, the total earthwork required increases.

Site-specific construction costs are greatly influenced by existing terrain features. Valleys, pits, hillsides, quarries, and other aspects of a particular site influence construction of dikes and dams. Ease of excavation is also a factor. Where excavation is relatively easy, the pond can be partly dug below grade and the material removed can be used for dike construction. Where the subsoil is rock or hardpan, material for dike construction may have to be acquired elsewhere, or blasting may be required.

Earthmoving costs can range from low values of less than $1/cu yd to over $10/cu yd where blasting is required (11, 12, 13). Since most of the construction cost for an unlined pond involves earthwork, costs for ponds can vary widely.

A final major component of ponds which significantly affects costs is the pond liner. Concrete liners can be used for small ponds of less than 1 acre. Clay linings or plastic linings can be used for larger ponds. Capital investment for a plastic lined pond may cost approximately 75 percent more than an unlined pond.

3.3 Land Treatment

Description

Land treatment is the application of wastewater directly to land areas. The wastewater is distributed over the land surface in basins or furrows by sprinkling or by overland flow. Waste materials are removed by filtration, adsorption, ion exchange, biological action, and plant uptake as the wastewater infiltrates the soil and/or passes over the surface.

The wastewater is usually pretreated by either primary treatment or by a combination of primary and secondary treatment. The pretreatment requirements depend on the raw wastewater composition and location of the land treatment area relative to the waste source and on the use classification of the agricultural crops that might be grown on the treatment land (4). After pretreatment, wastewater is piped to on-site storage tanks or basins and periodically discharged to the land treatment distribution system.

When the wastewater is permitted to percolate extensively through the soil, the treatment is referred to as infiltration. Infiltration treatment can be either slow-rate or rapid-rate. In slow-rate treatment, the wastewater is usually applied by sprinklers to moderately permeable soils. Site specific factors determine whether the sprinkler system is hand moved, mechanically moved, or permanently set (4). Rapid rate treatment is applied to deep and highly permeable soils such as sands or sandy loam. The water feeds from shallow basins formed by dikes constructed from surface soil. The underlying sandy soil within the basin then acts as a filtration medium. The wastewater distribution system continually supplies water to maintain a constant level in the basin during the application period, which may range from several hours to weeks.

Infiltration treatment systems can be designed to include underdrainage. Underdrainage consists of a network of drainage pipe buried beneath the land surface to recover the effluent, control groundwater contamination, or minimize the horizontal subsurface flow of leachate to adjoining property. The underdrainage network is usually intercepted by a collection ditch. Water can be recovered from the collection ditch for reuse or discharge to a receiving water body.

When wastewater does not percolate extensively, the procedure is referred to as overland flow. In this case, the wastewater is applied to the land at the upper end of a slope and allowed to run across the vegetated land surface into runoff ditches. Water quality is improved by physical, chemical, and biological interaction with a relatively impermeable surface soil layer.

A conceptual representation of land treatment systems is shown in Figure 3-10.

Land treatment can be used for municipal and certain industrial wastewaters. Preferably, these wastewaters should contain plant nutrients and be free from toxic materials. Municipal wastewater treatment by these methods produces effluents with BOD_5 and suspended solids concentrations both well below the 30 mg/l criterion of conventional secondary treatment.

Design Basis and Costs (4,6)

The potential for design variations in land treatment systems is so great that only general considerations are discussed here. These are consistent with the costs presented in one reference (4). The design basis for costs in another publication was not available (6). A summary of the design parameters is provided in Table 3-6. Costs are presented only for infiltration systems. Costs for other systems fall between the extremes of slow rate infiltration with and without underdrainage.

Sprinkler systems are used for slow rate infiltration. Costs include the use of bulldozer-type equipment for site clearing of brush and a few trees. Spray system specifications are as indicated in Table 3-6. A solid set or center pivot spray system is included. The design includes a 75-day wastewater storage reservoir with a distribution pumping station, standby pumps built into the dike of the reservoir, continuously cleaned water screens, all controls, and electrical work. Underdrainage is included for the category of slow-rate infiltration. The system does not include costs for pretreatment, monitoring wells, or transmission of water to and from the treatment facility.

Table 3-6. Design Parameters for Land Treatment (4)

Parameter	Units	Slow Rate Infiltration	
		Range	Design Value[a]
Field area required	acres/mgd	56-560	N/A[b]
Application rate	ft/yr	2-20	10
BOD_5 loading	lb/acre/d	0.2-5	N/A
Soil depth	ft	2-5	N/A
Soil permeability	in./h	0.06-2.0	N/A
Underdrain depth	ft	4-10	With and without
Spacing	ft	50-500	N/A
Application method	-	Sprinkling	Sprinkling

[a] N/A - Not available in cost Reference 4. Application rate was the only basis given for published cost data.

[b] The field area used for estimating the cost of land was 157 acres/mgd.

Total capital investment costs are plotted against design flow rate in millions of gallons per day in Figure 3-11. Several cost curves reflect the major variations in land treatment systems discussed above. The highest curve is for a slow-rate, sprinkler-fed, underdrained system. The lowest curve is for a slow-rate, gravity-fed system without underdrains. Costs for rapid and overland systems fall between

Figure 3-10. Land treatment system for wastewater

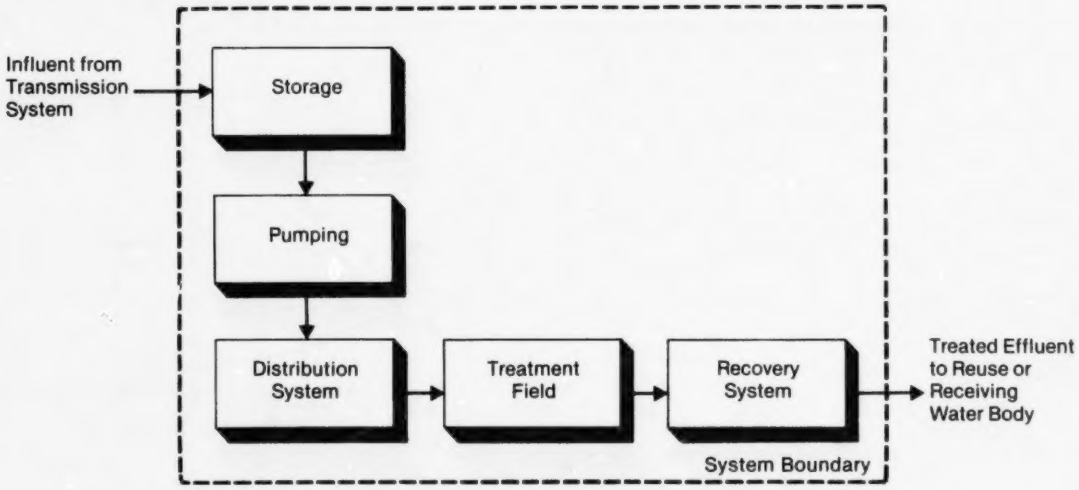
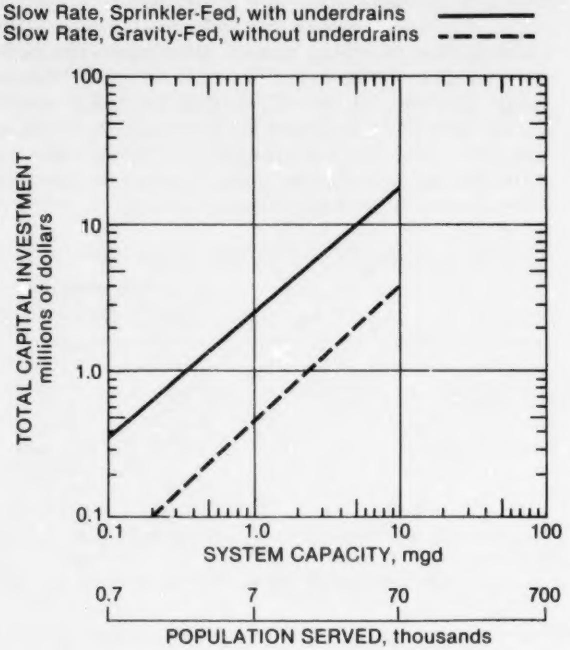

Influent from Transmission System
Storage
Pumping
Distribution System
Treatment Field
Recovery System
Treated Effluent to Reuse or Receiving Water Body
System Boundary

Figure 3-11. **Land treatment for wastewater treatment - Total capital investment (March, 1980 dollars).**



these two curves. Design details for the actual systems were not published, but costs reported for actual installations fell between these curves (6).

Operating costs expressed as net annual operating expenses are given in Figure 3-12. Unit annualized costs are given in Figure 3-13.

Major Variables Affecting Costs

It is highly speculative to generalize land treatment costs because local conditions have such a significant impact. The influence of some specific major variables can be discussed, however, to illustrate their relative significance for most systems.

The major factors in costs for land treatment are (11):

- Land Costs.
- Wastewater transmission costs to site.
- Site development costs including:
  - Relocation costs.
  - Land preparation.
  - Surface runoff control.
  - Subsurface drainage.
  - Distribution and irrigation.
  - Storage lagoons.
  - Pretreatment.

Land costs are highly variable, but would probably be relatively low in any area where land treatment would be considered as a viable treatment option.

Figure 3-12. **Land treatment for wastewater treatment - Net annual operating expense (March, 1980 dollars).**

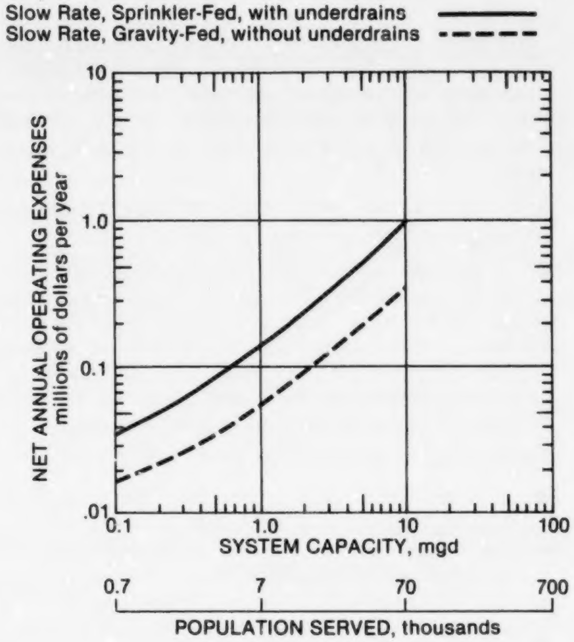
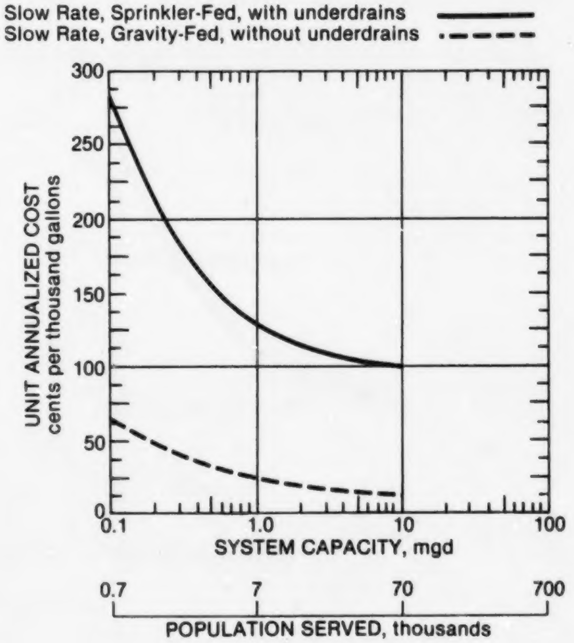


Figure 3-13. **Land treatment for wastewater treatment - Unit annualized cost (March, 1980 dollars).**



Costs of transmitting the wastewater to the site are highly variable and depend on the distance between the land treatment areas and the wastewater source. The transmission system is beyond the scope of the design considered in this report.

Site development costs are known to have varied by a factor of 36 for different locations (12). These costs include the various subcomponents listed above.

Culp, et al. (12) present a table summarizing the effect of various conditions on spray irrigation land treatment costs per acre. These conditions are classified as very favorable, moderately favorable, and unfavorable. Exclusive of system component, transmission, storage, or pretreatment costs, the ratio of costs with unfavorable to favorable treatment conditions was about 5 to 1.

As with capital investment costs, operating costs can vary widely. Labor, material, and energy requirements reflect site-specific factors. Further discussion of land treatment costs is given by Reed, et al.(13).

3.4 Phosphorus Removal by Chemical Addition

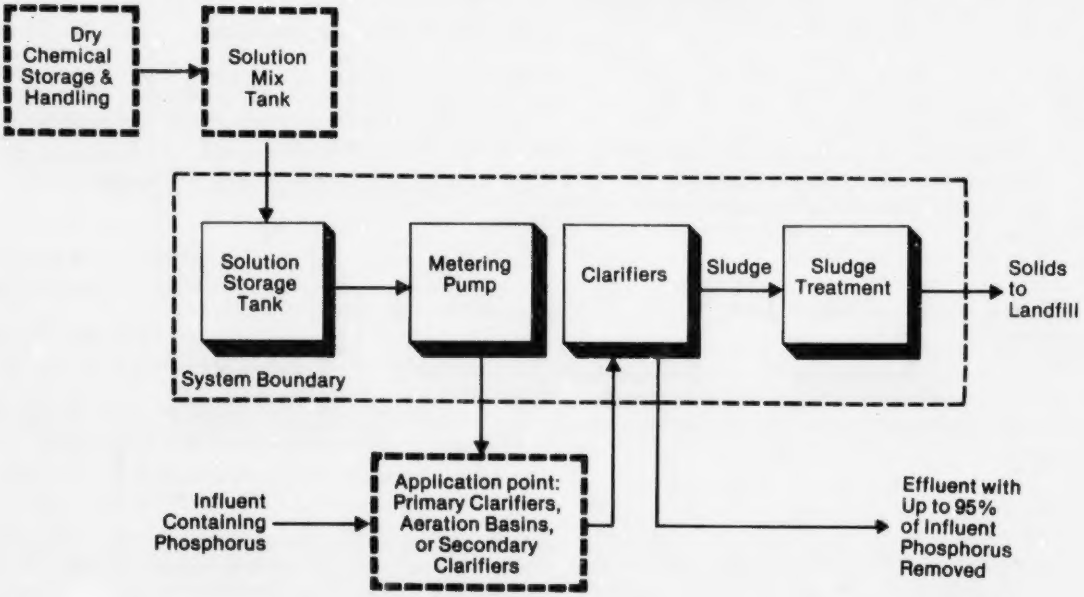
Description

Phosphorus is found in certain wastewaters as soluble chemical compounds and can be present in some suspended solids. Chemical addition is used to remove phosphorus in both the dissolved and suspended forms. Chemicals react with the dissolved phosphorus compounds (usually phosphates) to precipitate solids which can be removed by subsequent sedimentation in standard clarifiers. Three primary chemicals employed are lime, alum, and ferric chloride. The chemicals also help to coagulate suspended phosphorus solids (either precipitated from solution by chemical addition or originally present in the influent), thus facilitating their removal via sedimentation. Phosphorus removal efficiencies as high as 95 percent can be achieved using the chemical addition process (5).

The treatment chemicals can be added separately or in combination at various points in the overall treatment process. Possible points of addition include the primary clarifiers, activated sludge vessels, or secondary clarifiers in a conventional secondary or advanced wastewater treatment plant. These chemical additions are sometimes accompanied by the addition of polymers to further improve coagulation and sedimentation of solids.

Alum and ferric chloride may be purchased in dry form and prepared as solutions on-site or they may be purchased directly as solutions. The solution is metered from a holding tank to the point of application in either case. Lime is obtained dry and slaked with water to form a solid-liquid slurry. The slurry is similarly metered to the point of application. Figure 3-14 illustrates the major features of typical chemical addition systems.

Figure 3-14. Phosphorus removal by chemical addition for wastewater treatment.


Dry Chemical Storage & Handling
Solution Mix Tank
Solution Storage Tank
Metering Pump
Clarifiers
Sludge
Sludge Treatment
Solids to Landfill
System Boundary
Influent Containing Phosphorus
Application point: Primary Clarifiers, Aeration Basins, or Secondary Clarifiers
Effluent with Up to 95% of Influent Phosphorus Removed

Chemical addition can be used to remove phosphorus in municipal wastewaters and in industrial wastewaters where there are no substances in the wastewater that would interfere with the physical-chemical phosphorus removal mechanism. Dosage depends on the characteristics of the wastewater being treated.

Chemical treatment for phosphorus removal can be an add-on to existing or new secondary treatment plants, or it can be incorporated into independent physical-chemical treatment and tertiary treatment schemes, such as an adjunct to granular media filtration.

Design Basis and Costs (4,6,14,15)

For an alum or ferric chloride liquid feed system, the storage tank is designed for a 15-day supply of alum solution (49 percent strength) or ferric chloride solution (40 percent strength). The metering pump and piping are sized for twice the operating capacity. A building is included to house all major equipment for systems greater than 1 mgd. Smaller systems are not housed.

When the chemicals are received and stored in dry bulk form, the system would include additional equipment for dry chemical storage and handling, an agitated mix tank for liquid solution or slurry makeup, and the components of the liquid feed system described above.

Dosages for a phosphorus removal system are determined by jar test on the waste being treated. To reduce an influent phosphorus level of about 10 mg/l to less than 3.0 mg/l, typical chemical dosages are (4, 14, 15):

Alum, mg/l	200
Ferric chloride, mg/l	100
Lime, mg/l	150

Total capital investment is presented in Figure 3-15 as a function of the wastewater treatment plant design capacity. Cost curves are shown for an alum wet chemical feed system at two dosage levels (expressed as mg/l of alum). These costs correspond to the design basis discussed above and are presented schematically in Figure 3-14.

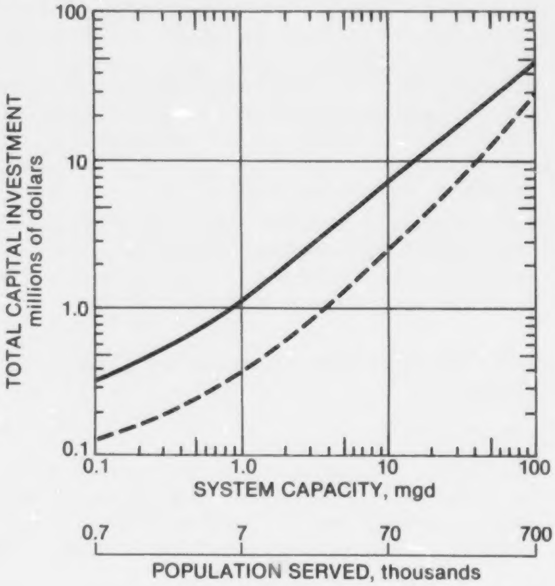
Net annual operating expenses are given in Figure 3-16 and unit annualized cost in Figure 3-17 as a function of the wastewater treatment plant design capacity. Again, multiple curves correspond to two dosage levels.

The upper curves correspond to dosages that would probably be required for phosphorus concentrations between 13 mg/l and 17 mg/l. The lower curves correspond to dosages for phosphorus concentrations in a range of about 5 to 9 mg/l. Costs for typical municipal wastewater phosphorus removal fall in the mid-range of the costs shown in the figures.

Figure 3-15. Phosphorus removal for wastewater treatment - Total capital investment (March, 1980 dollars).

High Alum Dosage, 300 mg/l ———
Low Alum Dosage, 100 mg/l - - - - -

Includes chemical feed system and incremental cost of clarifiers, sludge treatment, and sludge handling to accommodate increased sludge volume over plant not using phosphorus removal.



Major Variables Affecting Cost

The cost curves show the significant impact of dosage level on system cost. Dosage depends on influent phosphorus levels and pH which will vary for individual wastewater streams.

3.5 Nitrification (Separate-Stage)

Description

Nitrification is a biological process for ammonia removal in which ammonia is oxidized to nitrates and nitrites. Two uses of nitrification are:

- To convert nitrogen in ammonia to a form that can be removed in a downstream denitrification process.
- To convert nitrogen in ammonia to a form that does not have to be removed from the wastewater.

The cost information presented here is for separate-stage nitrification, which achieves the latter goal.

Single-stage nitrification occurs when operating conditions in the activated sludge process are adjusted to permit nitrogenous as well as carbonaceous oxidation to take place in the same aeration vessels. Separate-stage nitrification is a modified

Figure 3-16. Phosphorus removal for wastewater treatment - Net annual operating expenses (March, 1980 dollars).

High Alum Dosage, 300 mg/l ———
Low Alum Dosage, 100 mg/l – – – –

Includes chemical feed system and incremental cost of clarifiers, sludge treatment, and sludge handling to accommodate increased sludge volume over plant not using phosphorus removal.

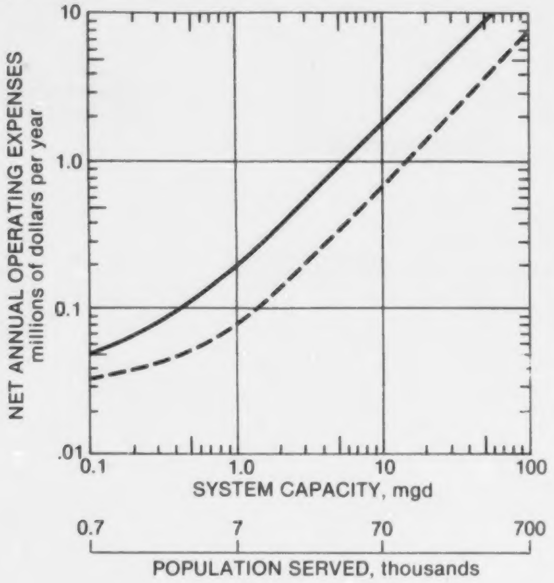
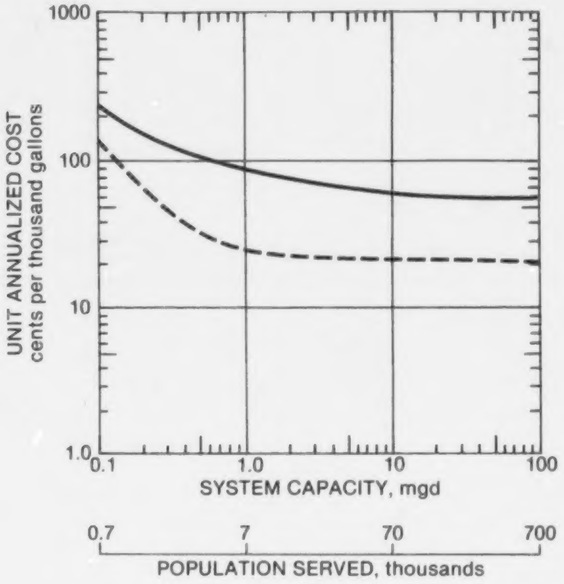


Figure 3-17. Phosphorus removal for wastewater treatment - Unit annualized cost (March, 1980 dollars).

High Alum Dosage, 300 mg/l ———
Low Alum Dosage, 100 mg/l – – – –

Includes chemical feed system and incremental cost of clarifiers, sludge treatment, and sludge handling to accommodate increased sludge volume over plant not using phosphorus removal.



single-stage process in which carbonaceous oxidation and nitrogenous oxidation occur in two separate aeration vessels and clarifier systems. The carbonaceous oxidation is the standard activated sludge process.

The nitrification system depicted in Figure 3-18 is designed to follow a high-rate activated sludge system. Since the process is pH sensitive, pH adjustment may be required as indicated. The incoming wastewater contains ammonia which is oxidized in two steps by autotrophic aerobic organisms called nitrifiers. Some organisms convert the ammonia to nitrite; others convert the nitrite to nitrate.

Design Basis and Costs (4,6,16)

Major design parameters and design values are given in Table 3-7. Secondary effluent is pumped to a plug flow (as opposed to completely mixed) nitrification tank constructed of concrete. The system also includes clarifiers (settling tanks) and all associated piping, pumps, electrical equipment, and instrumentation. Equipment for pH adjustment is not included but might be required for specific wastewater streams.

Total capital investment costs for a nitrification system with the design parameters shown in Table 3-7 are presented in Figure 3-19.

Net annual operating expenses are given in Figure 3-20, and unit annualized cost in Figure 3-21.

Major Variables Affecting Costs

As with other treatment systems with multiple components, equipment configuration can significantly affect costs. For example, one or several clarifiers could be used for the clarification step, and such a choice would probably alter capital investment costs by about 25 percent. A key design parameter that impacts tank volume, and therefore cost, is the mixed liquor volatile suspended solids (MLVSS) concentration required for the specific wastewater being treated. This requirement is roughly proportional to influent ammonia concentration. Over the typical operating range for this process, the reactor volume required for a high MLVSS concentration can be twice that for a low concentration. The capital cost of a high MLVSS concentration is about 50 percent more than that of a low concentration.

Operating expenses are also affected by the variables discussed above.

Figure 3-18. Separate-stage nitrification system for wastewater treatment.

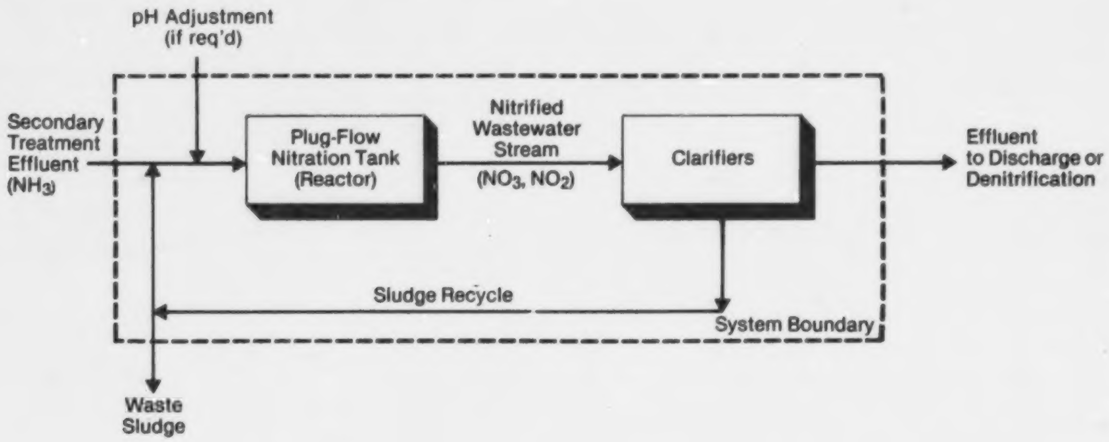

pH Adjustment
(if req'd)
Secondary
Treatment
Effluent
(NH_3)
Plug-Flow
Nitration Tank
(Reactor)
Nitrified
Wastewater
Stream
(NO_3, NO_2)
Clarifiers
Effluent
to Discharge or
Denitrification
Sludge Recycle
System Boundary
Waste
Sludge

Table 3-7. Design Parameters for Nitrification (Separate-Stage) (4)

Equipment	Parameters	Range of values	Design basis
Nitrification tanks (reaction)	Type	Plug-flow or complete mix	Plug-flow
	Waste loading (depends on temperature as well as MLVSS[a]), lb NH_3-Nitrogen lb/day/1000 ft^3	2.5-35	N/A[b]
	Operating temperature, °C	5-25	20
	MLVSS*, mg/l	1000-2000	N/A
	Detention time, hr	0.5 - 3.0	3
	Mean cell residence time, days	10 - 20	N/A
	Oxygen requirement, lb/lb NH_3-Nitrogen oxidized	N/A	4.5
	Dissolved oxygen (minimum), mg/l	2.0	N/A
	pH	7.2 - 8.5	8.4
Aeration equipment	Type	Mechanical or diffuser oxygen or air	Diffuser with air
Clarifiers	Overflow rate, gpd/ft^2	400-1000	600
	Solids loading, lb/d/ft^2	20-30	N/A
	Depth, ft	12-15	12
Sludge pumps	Sludge recycle	50%-100% operating average flow	Sized for twice operating flow rate 50%

[a] Mixed liquor volatile suspended solids.
[b] N/A - data not available.

Figure 3-19. Nitrification (separate-stage) for wastewater treatment - Total capital investment (March, 1980 dollars).

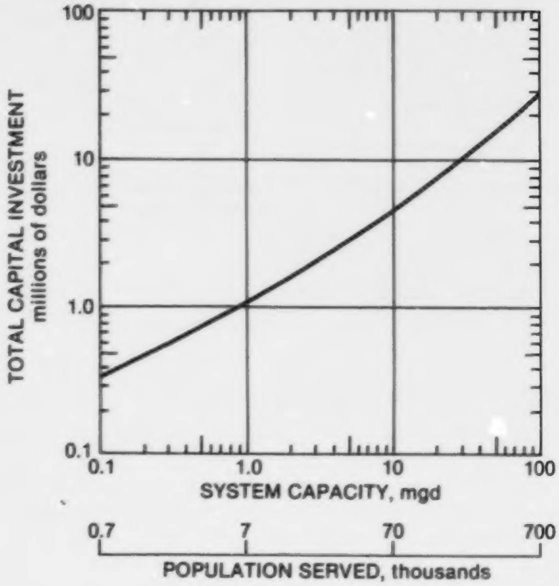


Figure 3-20. Nitrification (separate-stage) for wastewater treatment - Net annual operating expenses (March, 1980 dollars).

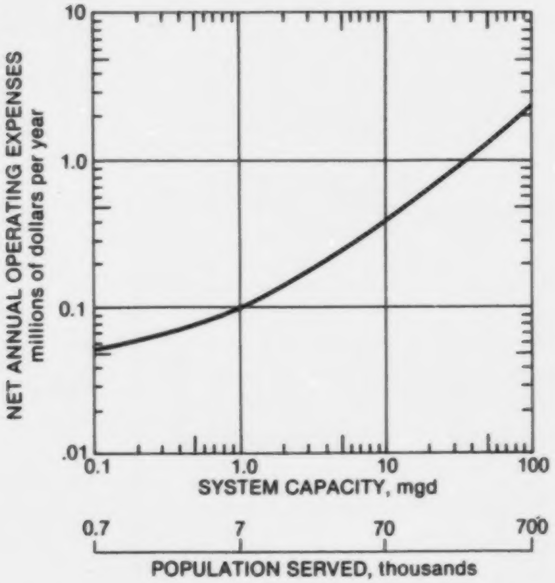
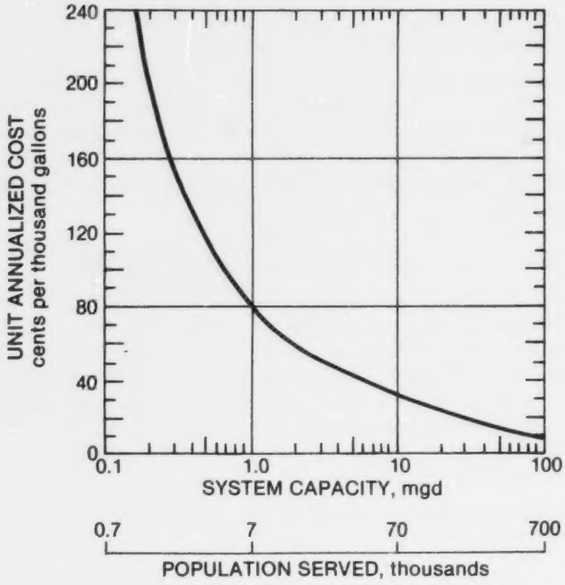


Figure 3-21. Nitrification (separate-stage) for wastewater treatment - Unit annualized cost (March, 1980 dollars).



3.6 Granular Media Filtration

Description

Granular media filtration is a treatment process used for the removal of suspended solids such as biological flocs or chemical flocs from secondary effluent. It can serve as a final polishing step or as a pretreatment for other processes where suspended solids interfere with performance such as granular activated carbon or reverse osmosis.

A granular media filter consists of either steel or concrete vessels containing the filter media. Steel vessels can be fed by either pressure flow or gravity flow. Concrete vessels are fed by gravity flow. Vessels may be subdivided into several compartments. The vessels contain specially designed support structures in the bottom to permit optimum drainage of the media.

Graded sand and anthracite coal are usually used as a filter media in dual media filters. In downflow filters, the sand is put in first and the coarser anthracite medium placed on top of the sand.

As wastewater flows through the filter bed, solids are deposited within the spaces between particles of the granular media. As solids build up, the pressure drop across the filter increases. At a predetermined pressure drop, the process flow is diverted to a parallel filter compartment or separate filter vessel while the first unit is cleaned by backwashing.

In the backwash cycle, some previously filtered water is pumped through the filter in the reverse flow direction to remove the deposited solids from the bed. Compressed air is sometimes introduced with the water to create a turbulent scouring action that dislodges compacted solid deposits.

Backwash water, containing the original wastewater solids at a much higher concentration than the original wastewater, is routed back to either existing clarifiers or vacuum filters for solids removal. The solids are disposed of with other treatment plant sludge.

Design Basis and Costs (4,6,15,17,18,19)

A complete filter system contains the major process elements shown in Figure 3-22. The filter unit generally consists of a vessel; the filter media; structures within the vessel to support the media; influent pumping and distribution devices; effluent pumps; and a backwash system of pumps, piping, and storage tanks. In addition to equipment, a building to house the filter system is included.

Design parameters and values upon which cost data is based are given in Table 3-8.

Total capital investment as a function of filter system capacity in millions of gallons per day is given in Figure 3-23. Figure 3-24 presents net annual operating expenses and Figure 3-25, unit annualized cost plotted against filter system capacity in millions of gallons per day.

Table 3-8. Design Parameters for Granular Media Filtration (4,16,18,19,20)

	Design Value	Typical Range
Hydraulic loading, gal./min/ft^2	4	2-8
Run length, hr	12	8-48
Backwash cycle time, min	15	—
Backwash hydraulic loading, gal./min/ft^2	15	15-25
Pump specifications:		
Type:	Centrifugal	
TDH, ft (overall)	14	
Efficiency, %	65	
Backwash holding tank:	Capacity for two backwash cycles	—
Bed depth, ft	—	2 to 4
Media depth ratio (sand to anthracite):	—	1:1 to 4:1
Air scour rate, scfm/ft^2	—	3 to 5
Terminal head loss, ft	—	6 to 15

Major Variables Affecting Costs

Capital investment for granular media filtration systems is sensitive to the hydraulic loading rate (gpm/ft^2) which in turn is determined by influent solids concentrations. Different design rates change filter vessel cross-sectional area and hence cost. Within the normal operating range, cost differences due to changes in loading are about 25 percent for the total system. The choice between steel or concrete vessels also affects costs. Concrete becomes more economical for very large systems.

Operating expenses are most sensitive to pump power requirements and frequency of backwashing.

Figure 3-22. Granular media filter system for wastewater treatment.

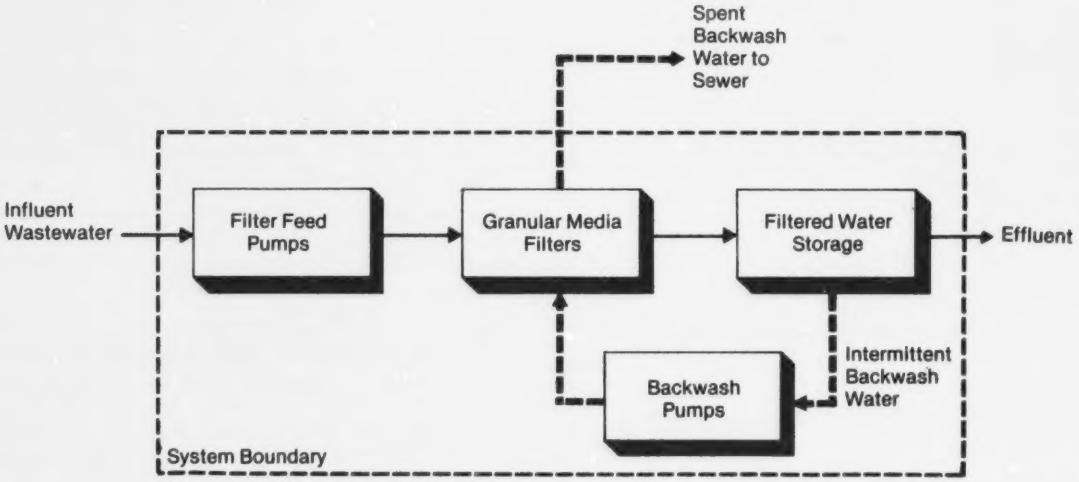


Figure 3-23. Granular media filtration for wastewater treatment - Total capital investment (March, 1980 dollars).

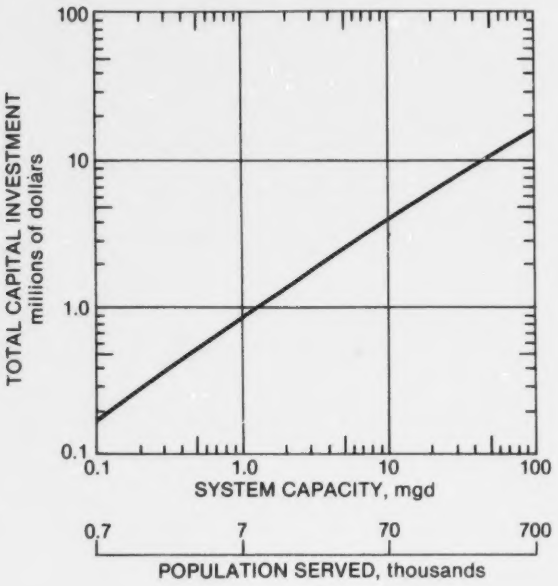


Figure 3-24. Granular media filtration for wastewater treatment - Net annual operating expenses (March, 1980 dollars).

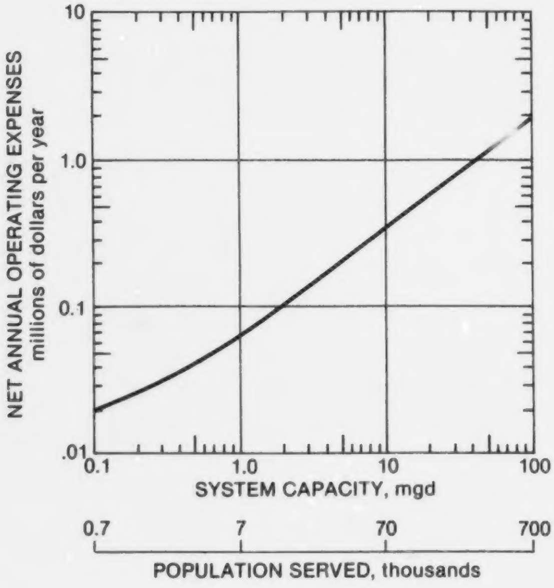
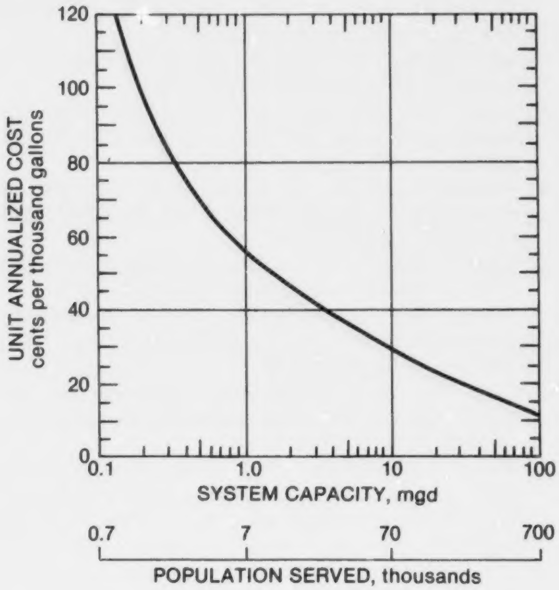


Figure 3-25. Granular media filtration for wastewater treatment - Unit annualized cost (March, 1980 dollars).



Power requirements depend on system design and head loss through the filter bed. Both system head loss and backwash frequency are functions of hydraulic loading, filter media characteristics, influent solids concentration, and solids characteristics.

References - Section 3

1. U.S. Environmental Protection Agency, Cincinnati, OH. Research Summary. Industrial Wastewater, EPA-600/8-80-026, June 1980.
2. U.S. Environmental Protection Agency, Office of Water and Waste Management, Washington, DC. Cost Estimates for Construction of Publicly-Owned Wastewater Treatment Facilities, Summaries of Technical Data, EPA-430/9-76-011, February 10, 1977.
3. The Cost of Clean Air and Clean Water. Senate Document No. 96-38. U.S. Government Printing Office, Washington, DC, 1979.
4. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009. U.S. Environmental Protection Agency, Office of Water and Waste Management, Washington, DC, February 1980, 260 pp.
5. Smith, C.V. and D. DiGregorio. Advanced Wastewater Treatment in Chemical Engineering, Deskbook Issue, April 27, 1970. pp. 71-74

6. Sage Murphy Associates. Construction Costs for Municipal Wastewater Treatment Plants, 1973-1978, EPA-430/9-80-003. U.S. Environmental Protection Agency, Office of Water and Waste Management, Washington, DC.
7. Metcalf and Eddy, Inc. Wastewater Engineering. McGraw-Hill, New York, NY. 1972.
8. Azad, H.S. (Ed.) Industrial Wastewater Management Handbook. McGraw-Hill, New York, NY, 1976.
9. Metcalf and Eddy, Inc. Wastewater Engineering: Treatment, Disposal, Reuse. McGraw-Hill, New York, NY, 1979. pp. 583.
10. Eckenfelder, W.W. Industrial Water Pollution Control. McGraw-Hill, New York, NY, 1966.
11. Parker, C.L. Estimating the Cost of Wastewater Treatment Ponds. Pollution Engineering, 7(11): 32-37, November 1975.
12. Culp, R.L., et al. Handbook of Advanced Wastewater Treatment, Van Nostrand Reinhold Co., New York, NY, 1978.
13. Reed, S.C., et al. Cost of Land Treatment Systems. EPA-430/9-75-003, U.S. Environmental Protection Agency, Washington, DC, Office of Water and Waste Management, September 1979.
14. U.S. Environmental Protection Agency, Cincinnati, OH. Process Design Manual for Phosphorus Removal, EPA-625/1-76-001a(PB259150*).
15. U.S. Environmental Protection Agency, Cincinnati, OH. Process Design Manual for Suspended Solids Removal. EPA-625/1-75-003a (PB259147*), January 1975.
16. U.S. Environmental Protection Agency, Cincinnati, OH. Nitrification and Denitrification Facilities; Wastewater Treatment, EPA-625/4-73-004a (PB259447*), August 1973.
17. Gumerman, R.C., et al. (Culp/Wesner/Culp) Estimating Water Treatment Costs, Vol. 1. Summary, EPA-600/2-79-162a (PB80-139819*). U.S. Environmental Protection Agency, Cincinnati, OH, August 1979.
18. Gumerman, R.C., et al. (Culp/Wesner/Culp) Estimating Water Treatment Costs, Vol. 2. Cost Curves Applicable to 1-200 mgd Treatment Plants, EPA-600/2-79-162b (PB80-139827*). U.S. Environmental Protection Agency, Cincinnati, OH, August 1979.
19. U.S. Environmental Protection Agency, Cincinnati, OH. Process Design Manual for Upgrading Existing Wastewater Treatment Plants, EPA-625/1-74-004a (PB259148*), October 1974.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

Section 4
Particulate Matter Collection

This section reports total capital investment, net annual operating expenses, and unit annualized costs for common systems used to remove particulate matter from gas streams. Particulate matter is one of the designated air pollutants recognized by the Clean Air Act that is regulated at the federal, state, and in some cases the local level. Particulate matter control is expected to continue to account for a significant fraction of the total expenditures for air pollution abatement. Between 1976 and 1982, for example, roughly half of the total investment expenditures for air pollution control are estimated to be for the purchase and installation of particulate matter control equipment (1).

For 1978, the total nationwide emissions of particulate matter were estimated at about 14 million tons, with 4.2 million tons from stationary combustion sources, 6.8 million from industrial processes, and the remainder from transportation and other sources (2). Fuel combustion emissions accounted for 33 percent of all particulate emissions, and electric utilities accounted for more than half of the fuel combustion emissions. The crushed stone, sand, and gravel industries accounted for 26 percent of the total particulate emissions (3). Although the current emissions profile probably differs somewhat from the 1978 profile, the data illustrate the relative magnitude of contributions from various industries. The total amount of particulate matter controlled by industry, excluding utilities, has been reported as about 40 million tons in 1978 (4).

Particulate matter emissions vary in particle size from submicron particles (less than 1 μm in diameter) to particles greater than 200 μm in diameter. Larger particles that comprise the greatest mass fraction of the emissions are the easiest to collect. Smaller inhalable particles (less than 15 μm) and respirable particles (less than 3 μm) are more difficult and costly to capture.

The four systems considered in this report are: mechanical collectors, electrostatic precipitators (ESP's), fabric filters, and wet scrubbers. Collection efficiencies and ranges of costs associated with these systems are compared in Table 4-1. Although other processes are available for particulate matter capture, this report considers only those systems that are widely used and which have been demonstrated to be capable of achieving removal efficiencies of at least 90 percent. Mechanical collectors (cyclones and multitube cyclones) have been used extensively to control particulate matter emissions, but their control efficiencies of 50 to 90 percent are lower than required by the more restrictive regulations for most applications. Mechanical collectors are often used for preliminary treatment in combination with other control systems. Multitube cyclones, which are more efficient than single tube cyclones, may find application as final particle collection devices in some situations.

The systems considered in this report can be categorized by the form in which the captured particulate material is removed from the collection

Table 4-1. Comparison of Major Particulate Collection Systems

Control system	Overall efficiency percent	Air flow rate range[a] acfm	Unit total capital investment dollars/acfm	Unit annualized cost[b] cents/acf
Multitube cyclone	50-90	10,000-1,000,000+	3-7	0.0004-0.0006
Electrostatic precipitator	80-99.5+[c]	10,000-1,000,000+	18-24	0.0006-0.006
Fabric filter	95-99.9[d]	10,000-1,000,000+	10-23	0.001-0.003
Wet scrubber (venturi)	75-99+	1,000-100,000+	6-12[e]	0.0007-0.004[e]

[a] Conversion factors to express air flow rate in other capacity units: 10^6 Btu heat input = 412 acfm; MWe output = 3200 acfm. The ranges of flow rates reflect the ranges of flow rates for which data were available for use in this report.

[b] Unit annualized costs are based on cost estimates presented in this report. The unit annualized cost accounts for all annual cash expenses and capital charges per unit of capacity.

[c] Most ESP's sold today are designed for 98 to 99.5% collection efficiency.

[d] Fabric filter collection efficiency is normally above 99.5%.

[e] The range for these costs is 10,000 to 100,000 acfm.

device. Wet scrubbers remove particles by contacting the gas stream with liquid, usually water, resulting in a slurry, or sludge, while cyclones, fabric filters, and ESP's remove particles directly from the gas in a dry form. Although some ESP's collect wet aerosols and fumes such as low-strength sulfuric acid mist or oil, these ESP's are not considered in this report because they are not widely used. Some advantages and disadvantages of wet and dry particulate matter removal systems are summarized in Table 4-2.

The overall performance of a given particulate matter collection system depends to a large extent on the specific characteristics of the particulate matter, particle size distribution, and the inlet gas stream particulate matter concentration. Figure 4-1 illustrates a typical relationship between removal efficiency and particle size. Fabric filters and ESP's exhibit high collection efficiencies for particles smaller than 3 μm. But as shown in Figure 4-1, wet scrubber collection efficiencies decrease significantly for smaller particles. Fabric filters generally offer the greatest potential for removing submicron particles or fines. Electrostatic precipitators are somewhat less efficient in removing fines, while high efficiency cyclones are the least efficient for these very small particles.

Table 4-2. Comparison of Wet and Dry Collection Systems

	Dry collection systems	Wet collection systems
Advantages	(1) Collected particulate matter weighs less than wet collected particulate matter and is therefore less costly to dispose of	(1) Ability to collect mists and aerosols
	(2) Usually requires less treatment for disposal	(2) Ability to collect gaseous pollutants in addition to particulate matter
	(3) Allows for particulate matter recovery in some cases	
	(4) Usuallv no serious corrosion problem	
Disadvantages	(1) Inability to collect mists and hygroscopic or caking materials	(1) May require wastewater and sludge treatment for disposal
	(2) Potential to create dust emissions in handling collected particulate matter	(2) Potential to discharge droplets of scrubbing liquor with entrained contaminants to the atmosphere

The performance of most fabric filter systems is not as strongly affected by changes in the inlet gas particulate matter concentration, as with ESP's, wet scrubbers, and mechanical collectors. With the latter devices, the outlet particulate matter concentration is directly related to the inlet particulate matter concentration.

Table 4-3 illustrates the effect of inlet concentration and emission concentration on required collection efficiency. The specified controlled emission concentrations represent typical new source performance standards and concentrations required by typical state regulations for existing sources. This table illustrates the levels of collection efficiencies for which most particulate matter systems must be designed.

Figure 4-1. Illustration of collection efficiency versus particle diameter (5).

A = Fabric Filter
B = Hotside ESP
C = Coldside ESP
D = Venturi Scrubber, ΔP = 100 in. H_2O
E = Venturi Scrubber, ΔP = 20 in. H_2O
F = Venturi Scrubber, ΔP = 10 in. H_2O
G = Multitube Cyclone
H = High Efficiency Single-tube Cyclone

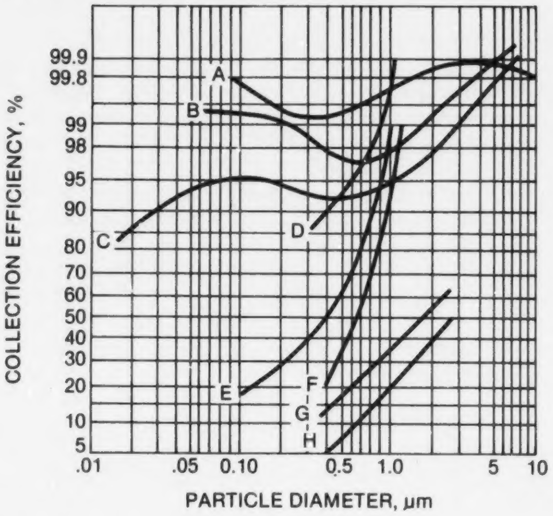


Table 4-3. Required Collection Efficiency for Typical Uncontrolled and Hypothetical Controlled Particulate Matter Concentrations

Inlet gas stream uncontrolled particulate matter concentration gr/acf[a]	Outlet gas stream controlled particulate matter concentration gr/acf[a,b]	Required control efficiency %
0.5	0.01	98.00
	0.05	90.00
	0.20	60.00
2.0	0.01	99.50
	0.05	97.50
	0.20	90.00
5.0	0.01	99.80
	0.05	99.00
	0.20	96.00
20.0	0.01	99.95
	0.05	99.75
	0.20	99.00

[a] gr/acf = grains per actual cubic foot. This is a common unit for expressed particulate matter concentrations. There are 7,000 grains to 1 pound.

[b] Outlet concentrations required by regulations for some sources include: grain elevators, 0.01 gr/acf; metals industry, 0.022 gr/acf; utility coal-fired boilers, 0.03 gr/acf.

Individual technologies and costs are described in the following sections.

4.1 Multitube Cyclones

Description

Cyclones use the principle of centrifugal separation to collect particulate matter in a dry form. There are two basic types of cyclones: single-tube cyclones and multitube cyclones. Multitube cyclones consist of a number of individual small diameter conically tapered tubes arranged in a common housing and operated in parallel. Spin vanes in each vertical tube of the multitube cyclone impart a high rotational velocity to the entering gas stream. As the gas stream spirals downward through the tubes, centrifugal forces impel the suspended particles toward the walls of the tubes. The particles fall from the open bottoms of the tubes into collection hoppers when the gas flow makes a sharp directional change upward at the bottom of each tube. From the collection hoppers the dust particles are transfered to storage, and ultimately, to disposal or recycle. The cleaned gas exits through the top of each tube into the outlet plenum and then to the stack.

Compared to single-tube cyclones, multitube cyclones achieve greater particle removal efficiencies without significantly increasing resistance to gas flow. Multitube cyclones may thus be used to meet particulate emission control requirements in some applications. However, the less efficient single-tube cyclones are usually unsuitable for this purpose and are thus not discussed in this report.

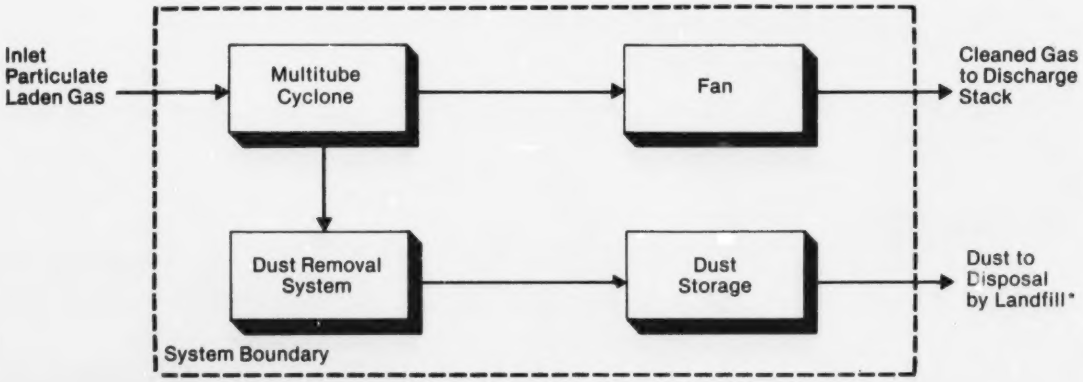
Complete multitube cyclone systems include the major components shown in Figure 4-2.

The efficiency of multitube cyclones depends mainly on the inlet gas velocity, the diameter and length of individual tubes, and most importantly, the range of particle sizes in the entering gas stream. Higher inlet gas velocities, smaller tube diameters, and longer tube lengths increase particle removal efficiency but also increase resistance to gas flow. As was shown in Figure 4-1, smaller particles are collected less efficiently than larger particles. The overall collection efficiency thus depends on the relative proportions of small and large particles. Efficiencies achieved in various cyclone applications for emissions control vary from about 55 to 95 percent. Applications include removal of fly ash from coal-fired boilers and dust control in minerals processing.

Design Basis and Costs

The principal variables in the design of multitube cyclone systems are: gas flow rate, inlet particulate matter concentrations, particle size distribution, desired particle removal efficiency, and potential need for corrosion or erosion resistant materials. The design characteristics (such as tube diameter and inlet gas velocity) of the cyclone depend on all of the listed design variables. The size of the associated ductwork depends on the gas flow rate. Fan size depends on the gas flow rate and the system pressure drop (which depends on resistance to the gas flow). The particulate matter inlet concentration and the

Figure 4-2. Multitube cyclone system for Particulate matter collection.


Inlet Particulate Laden Gas
Multitube Cyclone
Fan
Cleaned Gas to Discharge Stack
Dust Removal System
Dust Storage
Dust to Disposal by Landfill*
System Boundary

*If the dust is valuable, it may be recovered as product or for recycle. The costs in this report assume that the dust is disposed of as landfill.

particle removal efficiency determine the quantity of dust removed and thus determine the size and type of dust removal system.

Table 4-4 presents the design parameters for the multitube cyclone systems used to develop the cost curves presented in this section. The cyclone design characteristics and inlet particulate matter concentration typify multitube cyclone applications on coal-fired boilers.

Total capital investment data are presented in Figure 4-3. Net annual operating expenses are shown in Figure 4-4, and unit annualized costs are shown in Figure 4-5.

The capital investment data were developed by updating the correlating cost data from the following references: multitube cyclone costs (7); fan and motor costs (8); ducting costs (9); and dust removal costs (10). Net annual operating expenses and unit annualized costs were developed from data in Reference 9.

Major Variables Affecting Costs

The investment cost of the multitube cyclone itself is usually less than half of the capital investment of the total multitube cyclone system. In the system considered in this report, the multitube cyclone cost is only 17 to 43 percent of the system cost, with a proportionately greater share for the larger systems. Significant variations in the multitube cyclone characteristics (such as tube diameter) thus may not significantly affect system costs.

The fan and motor cost comprises 8 to 22 percent of the investment costs for the systems considered in this report. For the larger systems, the fan and motor cost is the largest element of the investment.

Table 4-4. Multitube Cyclone Design Parameters

Parameters	Design basis	Typical range
Gas flow rate, acfm	10,000; 100,000; 1,000,000	a
Inlet loading, gr/acf	2.0	b
Overall particle Removal Efficiency, %	85	55-99+^{c}
Pressure drop, in. H_2O	3	2-6^c
Operating pressure	Atmospheric	b
Tube diameter, in.	9	6-24^c
Tube length, ft	b	b
Materials of construction	Carbon Steel	Carbon Steel
Dust removal and storage		
-1,000 and 10,000 acfm	Dumpster	Site-specific
-100,000 and 1,000,000 acfm	pneumatic conveying/ storage silo	Site-specific
Dust disposal	Landfill	Landfill, bonding, or recycle
Ducting, ft	100	Site-specific
Operating factor, %	70	Site-specific
Equipment life, years	20	b

[a] Up to about 100,000 acfm for single multitube units. Higher gas flows require multiple units.
[b] Unknown or not specified.
[c] Reference 6.

The cost of ducting ranges from about 12 to 46 percent of the investment costs for the systems considered in this report. For the smaller systems, the ducting cost is the largest element of the investment.

The cost of the ash removal and storage system ranges from about 19 to 60 percent of the investment costs, with the greater share for intermediate-sized systems.

The major contributors to net annual operating expenses are: dust disposal costs, electricity for the fan, labor, and maintenance costs. Electricity and dust disposal costs are a much more significant portion of total costs for the large systems than small systems. Electricity costs are up to about 20 percent of the net annual operating expenses. Dust disposal costs comprise 50 to 65 percent of the expenses. Thus, net annual operating expenses can be significantly reduced if ultimate disposal costs can be reduced or credits can be taken for recovered useable material.

4.2 Electrostatic Precipitators

Description

Electrostatic precipitators (ESP's) are used to remove particulate matter from waste gases in a variety of industrial applications. Industries and emission

Figure 4-3. Multitube cyclone system for particulate matter collection - Total capital investment (March, 1980 dollars).

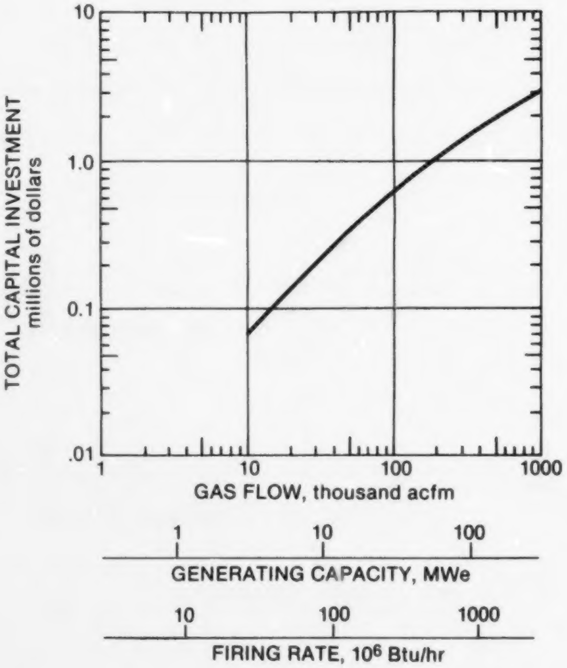


Figure 4-4. Multitube cyclone system for particulate matter collection - Net annual operating expenses (March, 1980 dollars).

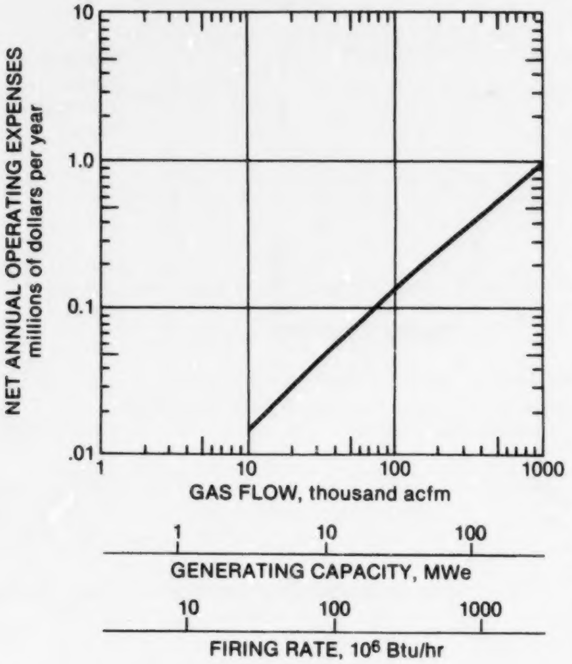
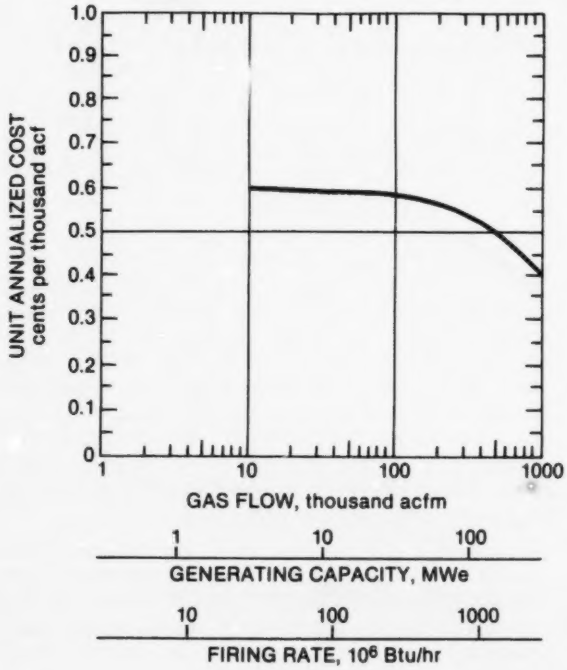


Figure 4-5. Multitube cyclone system for particulate matter collection - Unit annualized cost (March, 1980 dollars).



sources which commonly use ESP's for particulate matter collection are: aluminum industry (baking ovens, reduction cells, carbon plants, bauxite dryers, alumina calciners, and remelt furnaces), carbon black production, sulfuric acid recovery, asphalt blowing stills, phosphoric acid production, tar and oil recovery from waste or fuel gases, fluid-bed catalytic crackers, coal-fired boilers, pulp and paper industry, cement industry, gypsum industry, iron and steel industry (sintering and furnaces), municipal waste incinerators, glass manufacturing, phosphate rock crushing lime industry, and copper smelters. Particle removal efficiencies of greater than 99.9 percent have been achieved in some of these applications.

The collection efficiency of an ESP is primarily dependent on the characteristics of the particulates (particle size and electrical resistivity) and the amount of collection electrode plate surface area (SCA) used.

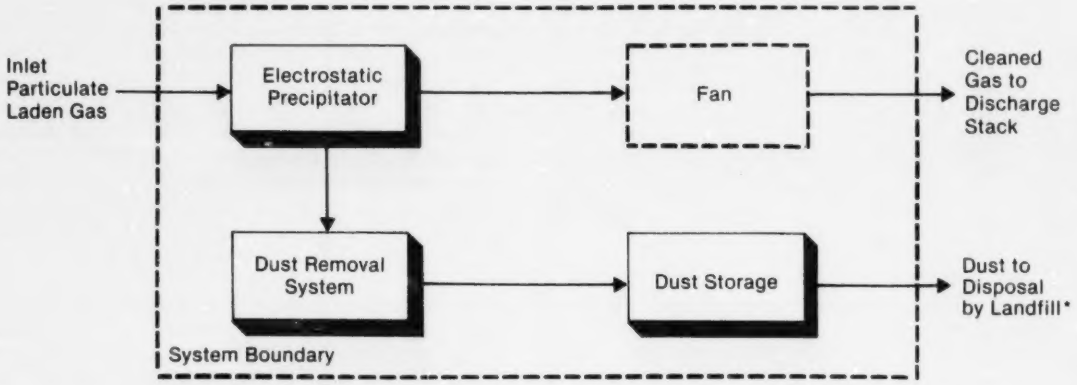
Figure 4-6 is a conceptual diagram of a typical ESP installation which shows four major components: the ESP, ducting, a fan, and a collected solids (dust) removal system. Because the pressure drop across the ESP itself is negligible relative to the total system pressure drop, in most ESP applications, fans are normally not included in the scope of an ESP system design and are not considered in the cost estimates presented here.

The ESP system shown in Figure 4-6 features dry particle removal on vertical plate electrodes. Particulate matter removal occurs in three steps: 1) suspended particles are given an electrical charge, 2) the charged particles migrate to a collecting electrode of opposite electrical charge, and 3) the collected particles are dislodged from the collecting electrodes by mechanical rapping of the electrodes. This rapping causes the particles to fall into hoppers underneath for subsequent disposal.

The complexity of the dust removal system associated with an ESP depends primarily on the size of the system. In small ESP systems dust is usually discharged directly from hoppers to dumpsters; the dust is ultimately disposed of as landfill. Large ESP systems typically feature either a pneumatic dust removal system that transfers dust to a storage silo before ultimate disposal, or wet sluicing of ash to wastewater treatment and ultimate disposal. The collected dust could also be recovered as product or recycled.

Gas precleaning with mechanical collectors can sometimes effectively reduce the size (and thus costs) of an ESP used for final particulate matter removal. The inlet gas stream may also be pretreated with water sprays and/or chemical additives to improve the electrical resistivity of the particles, thereby increasing the overall capture efficiency.

Figure 4-6. Electrostatic precipitator system for particulate matter collection.


Inlet Particulate Laden Gas
Electrostatic Precipitator
Fan
Cleaned Gas to Discharge Stack
Dust Removal System
Dust Storage
Dust to Disposal by Landfill*
System Boundary

*If the dust is valuable, it may be recovered as product or for recycle. The costs in this report assume that the dust is disposed of as landfill.

Design Basis and Costs

The scope of the ESP systems discussed in this section is illustrated in Figure 4-6. Equipment or operations shown by solid lines in Figure 4-6 are included in the scope of ESP systems. The fan shown by dashed lines is not included in the scope of ESP systems. This is because ESP pressure drop is relatively small, and the installation of an ESP on a source already requiring a fan will not result in a significant alteration in fan requirements.

The principal variables in the preliminary or conceptual design of ESP systems are: inlet particulate matter loading, desired removal efficiency, particle size distribution, particle electrical resistivity (affected by the gas composition and temperature as well as particle composition), gas flow rate and distribution, and materials of construction. The particle inlet loading, gas flow rate, and removal efficiency determine the quantity of dust removed and thus determine the size and type of dust removal system needed. The particle sizes and resistivity determine the ease with which a specific particulate matter precipitates and thus affects ESP design. As with other gas handling equipment, flow rate also affects ESP design and the size of associated ductwork. If the gas is corrosive, then higher-price corrosion-resistant materials must be incorporated into the system.

Classically, ESP's have been sized from the Deutsch-Anderson equation for preliminary or conceptual designs:

$$\eta = 1 - \exp\left[-\frac{A}{V}\omega\right] \qquad (4\text{-}1)$$

where η =desired removal efficiency, expressed as a fraction
A=collecting electrode plate area (the principal ESP design parameter), ft^2
V=gas volume flow rate, ft^3/s
ω =precipitation rate parameter, ft/s

The precipitation rate parameter, ω, is determined from experience for specific applications and accounts for differences in particle sizes and resistivity from one application to another. The physical significance of ω is to describe the average rate at which charged particles will migrate toward the collecting electrodes. Lower values of ω denote more difficult precipitation applications such as the precipitation of high resistivity fly ash from low sulfur coals. Table 4-5 shows the range of precipitation rate parameters encountered in typical ESP application.

Table 4-6 presents the design parameters for 12 case study ESP's used to develop the cost curves presented in this section. The only constant design variable for these ESP's is the inlet particle loading, which at 2.0 gr/acf, typifies flue gases from coal-fired boilers, lime kilns, and other sources. Higher loadings results in larger and more expensive ash disposal systems and also increase the cost of ultimate disposal. The potential impact of higher loadings is addressed in the discussion of major variables affecting costs. High loadings could also adversely affect the precipitation rate.

Figure 4-7 shows the total capital investment required for ESP's used in applications ranging from 100,000 to 1,000,000 acfm as a function of two major design variables: the ease of precipitation, and the required particle removal efficiency. In using Figure

Table 4-5. Precipitation Rate Parameters for Typical ESP Applications[a]

Emission source or industry category	Precipitation rate parameters, fps: Range	Precipitation rate parameters, fps: Average or typical	Difficulty of precipitation[h]
Coal-fired boilers	0.10-0.67[b,c,d,e]	0.35[f]	Difficult to easy
Pulp and paper industry	0.2-0.35[b,c]	0.25[b,d]	Average
Cement industry	0.19-0.45[c,d]	0.30[f]	Difficult to average
Gypsum industry	0.4-0.64[c,d]	0.52[f]	Average to easy
Iron and steel industry			
-sintering	0.08-0.4[c]	0.27[f]	Difficult to average
-open-hearth furnace	0.15-0.3[c]	0.21[f]	Difficult to average
-basic oxygen furnace	0.15-0.25[c]	0.20[f]	Difficult to average
-electric arc furnace	0.12-0.16[c]	0.14[f]	Difficult
-blast furnace	0.20-0.46[b]	0.37[f]	Average
-gray iron cupola	0.10-0.12[b,d]	0.10[b]	Difficult
Municipal waste incinerators	0.2-0.4[c]	0.3[f]	Average
Glass manufacturing	—	0.14[e]	Difficult
Phosphate rock crushing	—	0.35[e]	Average
Lime industry	0.17-0.25[e]	0.25[g]	Difficult to average
Copper smelters	0.12-0.14[c]	0.13[f]	Difficult
Other smelters	—	0.25[c,d]	Average
Petroleum cat cracking	0.12-0.18[e]	0.15[f]	Difficult

[a] The precipitation rate parameter has classically been used in the Deutsch-Anderson equation to predict ESP performance in various applications. Although it can be assumed constant some applications, the precipitation rate parameter will actually vary in given application as a function of collection efficiency. Use the of Deutsch-Anderson equation to design (size) highly efficient ESP's requires using a conservative (low) precipitation rate parameter. For example, the precipitation rate parameter for an ESP on one coal-fired boiler 0.43 fps for a 92% removal and 0.16 fps for a 99.5% removal. (Reference 1)

[b] Reference 11

[c] Reference 12

[d] Reference 13

[e] Reference 8

[f] Estimate

[g] Reference 14

[h] Difficult = Precipitation Rate Parameter <0.20 fps
Average = Precipitation Rate Parameter >0.20 fps, <0.50 fps
Easy = Precipitation Rate Parameter >0.50 fps

Table 4-6. Design Parameters for Model Electrostatic Precipitators[a]

Ease of precipitation	Precipitation rate parameter (ω)[b] fps	Gas flow rate acfm	PM removal efficiency %	Collecting electrode surface area[b] ft^2	PM removal rate lb/hr
1. Difficult	0.10	10,000	95	4,990	163
2. Difficult	0.10	10,000	99.9	11,500	171
3. Difficult	0.10	100,000	95	49,900	1,630
4. Difficult	0.10	100,000	99.9	115,000	1,710
5. Difficult	0.10	1,000,000	95	499,000	16,300
6. Difficult	0.10	1,000,000	99.9	1,150,000	17,100
7. Typical	0.30	10,000	95	1,660	163
8. Typical	0.30	10,000	99.9	3,840	171
9. Typical	0.30	100,000	95	16,600	1,630
10. Typical	0.30	100,000	99.9	38,400	1,710
11. Typical	0.30	1,000,000	95	166,000	16,300
12. Typical	0.30	1,000,000	99.9	384,000	17,100

[a] All 12 model ESP's have the following additional characteristics: (1) Inlet dust loading - 2.0 gr/acf, (2) Operating factor = 70%, (3) Equipment lifetime = 20 years, (4) Power demand = 3.5 W/ft^2 of collecting electrode area, (5) Duct length = 200 ft, and (6) Mild steel construction. The 10,000 acfm systems feature ash removal and storage in a dumpster. The larger systems feature pneumatic conveying of ash to silos for storage.

[b] The collecting electrode surface area is estimated from the Deutsch-Anderson equation:

$$\eta = 1 - \exp - \frac{A}{V} \omega$$

where η = PM removal efficiency
A = Collecting electrode surface area, ft^2
V = Gas flow rate, ft^3/sec
ω = Precipitation rate parameter, fps

Figure 4-7. Electrostatic precipitator system for particulate matter collection - Total capital investment (March, 1980 dollars).

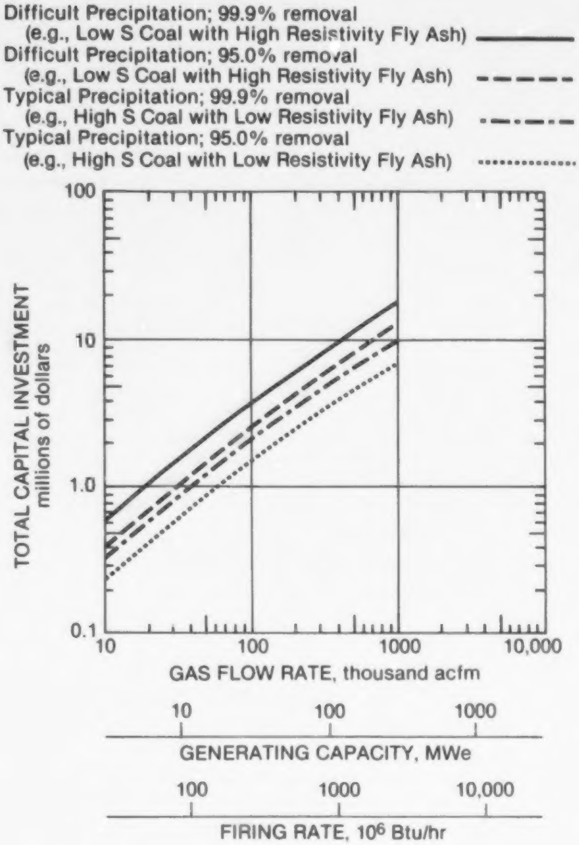
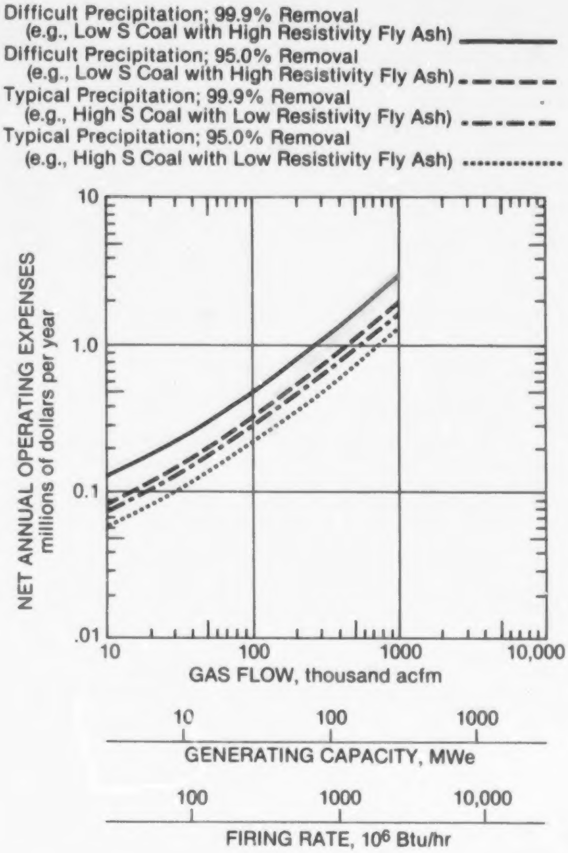


Figure 4-8. Electrostatic precipitator system for particulate matter collection - Net annual operating expenses (March, 1980 dollars).



4-7 (and Figures 4-8 and 4-9), the user should refer to Table 4-5 to determine the ease of precipitation (i.e., the precipitation rate parameter) for the application of interest.

Net annual operating expenses are shown in Figure 4-8, and unit annualized costs are presented in Figure 4-9.

The capital investment cost curves are developed from statistical correlations of study estimates of the various ESP subsystems from References 8, 10, 14, 15, and 16. The subsystems are the ESP, ducting, and a solids disposal system. The ducting and solids disposal subsystem costs were estimated using correlations based on the costs presented in two references (10, 15). The net annual operating expense curves are derived from References 10, 14, 15, and 16.

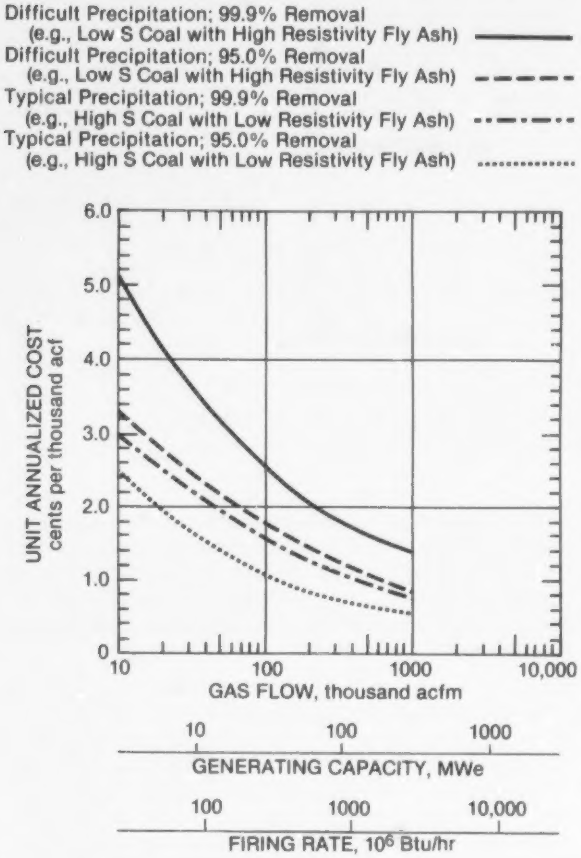
Major Variables Affecting Costs

As discussed previously, high particulate matter concentrations result in larger and more expensive dust collection systems and increase the cost of ultimate disposal. High concentrations may also adversely affect the precipitation rate and require a larger, more expensive ESP. A higher particle concentration that increases the capital investment of an ash removal system by 100 percent increases the total capital investment up to 25 percent depending on overall system size. A 100 percent increase in the cost of ultimate solids disposal can increase the net annual operating expenses by as much as 50 percent, again depending on overall system size. There is a relatively higher percentage impact on larger systems.

For some applications, a fraction of the recovered solids can be recycled. As illustrated above, net annual operating expenses can be significantly reduced if ultimately disposal costs can be reduced or credits taken for recovered usable material. Such evaluations must be made on a case-by-case basis.

Although the cost curves in Figures 4-7 through 4-9 are based mainly on data for ESP's applied to coal-

Figure 4-9. Electrostatic precipitator system for particulate matter collection - Unit annualized cost (March, 1980 dollars).



fired boilers, they are applicable to ESP's for other emission sources if the inlet grain loading and other design parameters are similar.

4.3 Fabric Filters

Description

Fabric filters are widely used to control dry particulate matter emissions in a variety of industrial applications, whenever dry bulk solids are processed, especially in processsing metals, minerals, and grains. Fabric filters are also becoming more widely used on coal-fired boilers.

Fabric filtration is the physical straining or sieving of particulate matter from gas streams. The gas stream passes through a fabric filter medium (usually in the shape of a bag). Particulate matter from the inlet gas deposits mainly on the surface of the filter bag, where a dust layer accumulates. Both the collection efficiency and the pressure drop across the bag surface increase as the dust layer builds up. When the pressure drop becomes excessive the bags are cleaned to remove collected solids which are then disposed of.

The collection efficiency of fabric filters is primarily dependent on the characteristics of the fabrics used and particulate matter (particle size distribution and cake porosity) and is not dependent to any noticeable extent on the amount of collection fabric surface area. However, the fabric surface area does affect pressure drop and hence the energy requirements.

Filters are generally cleaned in one of three ways. In shaker cleaning, the filter bags are oscillated by a small electric motor. The oscillation dislodges varying amounts of dust into a hopper depending on the shaking frequency and amplitude. In reverse air cleaning, a reverse air flow is used to collapse the bags, and fracture and dislodge the dust cake. Both shaker cleaning and reverse air cleaning require a sectionalized baghouse to permit cleaning of one section while other sections are functioning normally. The third cleaning method, pulse jet cleaning, does not require sectionalizing (with some dusts, sectionalization may be preferred for pulse jet cleaning). A short pulse of compressed air is delivered through nozzles at the bag exit and is directed toward the bottom of each bag. The primary pulse of air entrains a pulse of secondary air flow as it passes through the nozzles. The resulting pressure produces a shock that expands the bag and dislodges the surface dust layer.

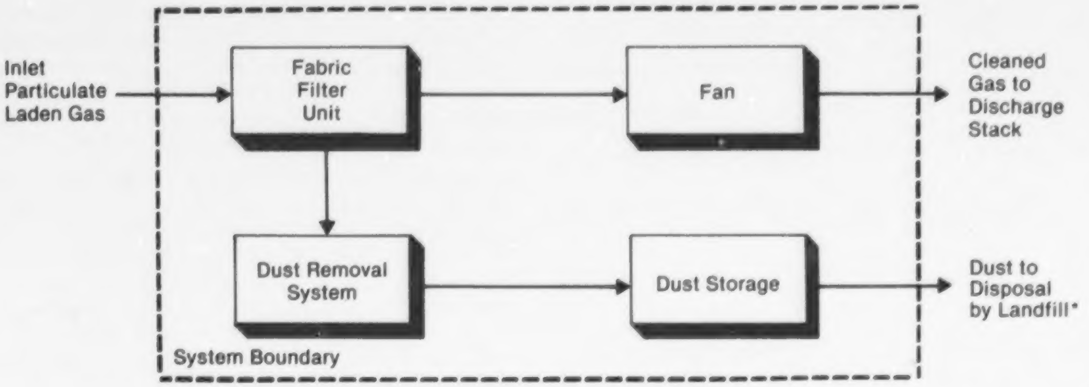
Although all three basic cleaning methods, singly or in combination, are used, only reverse air and pulse jet cleaning methods are discussed in this report. Reverse air cleaning appears to be preferred for larger applications while pulse jet cleaning appears to be preferred for smaller applications. Shaker cleaning is most suitable when the filtration medium will not degrade due to mechanical stresses.

Figure 4-10 is a conceptual diagram of a typical fabric filter installation. Fabric filter systems consist of four major components: a baghouse (containing the fabric filter bags), ducting, a booster fan, and a solids removal system.

The complexity of a fabric filter system depends to a large degree on the quantity of solids to be collected. Small fabric filter systems typically discharge collected solids directly from baghouse hoppers to dumpsters; the solids are ultimately disposed of by landfilling. Large fabric filter systems typically feature either a pneumatic dust removal system which transfers dust to a storage silo and ultimate disposal, or wet sluicing to water treatment and ultimate disposal. The collected dust could also be recovered as product or for recycle.

Fabric filter systems are sometimes preceded by gas pretreatment or conditioning processes. Precleaning is usually accomplished with mechanical collectors. Since most common filter fabrics are limited to temperatures below 550°F (with some fabrics limited

Figure 4-10. Fabric filter system for particulate matter collection.


Inlet Particulate Laden Gas
Fabric Filter Unit
Fan
Cleaned Gas to Discharge Stack
Dust Removal System
Dust Storage
Dust to Disposal by Landfill*
System Boundary

*If the dust is valuable, it may be recovered as product or for recycle. The costs in this report assume that the dust is disposed of as landfill.

to temperatures below 175°F), the gas stream can be cooled either directly by quenching the gas with water or by indirect cooling. The temperature must be maintained high enough, however, to prevent moisture condensation within the baghouse and on the filter surface. Moisture causes the collected particles to agglomerate and plug the bags.

Particulate matter removal efficiencies of greater than 99.9 percent are achieved in many applications, with particulate matter concentrations in the filtered gas usually less than 0.04 gr/ft^3. Fabric filters can effectively control fine particulates and usually require only moderate pressure drop across the control system. Fabric filter performance is relatively unaffected by flue gas composition, inlet particulate matter concentration, and particle composition and properties.

Design Basis and Costs

The principal variables in the design of fabric filter systems are: gas flow rate, bag cleaning method, filter bag surface area, inlet particulate matter concentration and size properties, potential need for corrosion resistant and heat-resistant materials, gas temperature, and gas moisture content. The required filter area for a specific application depends on the gas flow rate, the bag cleaning method, and characteristics of the dust. The sizes of associated ductwork and fan are also dependent on the gas flow rate. The particulate matter inlet concentration determines the quantity of dust removed and thus determines the size and type of solids removal system. High inlet particulate matter concentrations may also affect the filter area or require precleaning.

The most important design variables, the bag cleaning method and fabric filter area, are determined from experience in similar applications. Although either reverse air cleaning or pulse jet cleaning may be used for most applications, a specific cleaning method may be preferred. Pulse jet cleaning allows the use of smaller fabric filter areas than does reverse air cleaning.

For a specific application, bag material, and bag cleaning method, the required filter area is determined from the gas flow rate and the air-to-cloth ratio demonstrated by experience to be most suitable and economical. The air-to-cloth ratio is the ratio of the gas flow rate to the filter area and is typically expressed as actual cubic feet per minute per square foot (acfm/ft^2) or feet per minute (fpm). The air-to-cloth ratio for baghouses using reverse air cleaning averages 2 fpm with a range of 1.5 to 3.5 fpm. The air-to-cloth ratio for baghouses using pulse jet cleaning averages 6 fpm with a range of 4 to 15 fpm (13, 17).

Table 4-7 presents the design parameters for five case study fabric filter systems used to develop the cost curves presented in this report. The inlet particulate matter concentration of 2 gr/ft^3 assumed for these systems is representative of flue gases from coal-fired boilers (10, 11, 12, 14, 15). The potential cost impacts of higher or lower air-to-cloth ratios and inlet particulate matter loadings are addressed in the discussion of major variables affecting costs.

Total capital investment is presented in Figure 4-11. Net annual operating expenses are shown in Figure 4-12, and unit annualized cost estimates are shown in Figure 4-13.

The capital investment curves are derived from study estimates reported in References 8, 10, 14, and 16. The cost curves are developed from statistical correlations of the estimated costs of the various fabric filter subsystems: baghouse and filter bags, ducting, fan, and solids removal system. The ducting and solids disposal subsystem costs are based on correlations derived from costs for these subsystems in References 10 and 16.

Table 4-7. Design Parameters for Model Fabric Filter Systems[a]

Gas flow rate acfm	Particulate matter removal rate lb/hr
10,000	170
100,000	1,700
1,000,000	17,700

[a] All model fabric filter systems have the following additional characteristics: (1) Reverse-air air-to-cloth ratio = 2 fpm, (2) Pulse-jet air-to-cloth ratio = 6 fpm, (3) Inlet dust concentration = 2.0 gr/scf, (4) Outlet dust concentration = 0.02 gr/acf or less, (5) Operating factor = 70%, (6) Equipment lifetime = 20 years, (7) Bag lifetime = 2 years, (8) Duct length = 200 ft, and (9) Mild steel construction with insulation. The reverse-air systems are assumed to operate at high temperature (<550°F) and use fiber-glass bags. The 10,000 acfm system features ash removal and storage in a dumpster. The 100,000 and 1,000,000 acfm systems feature ash removal in a pneumatic conveying system and storage in a silo.

[b] The air-to-cloth ratio (A/C) and the gas flow rate (A) determine the required fabric filter area (C) since C = A/(A/C).

Figure 4-11. Fabric filter system for particulate matter collection - Total capital investment (March, 1980 dollars).

Curves based on reverse air cleaning with air-to-cloth ratio of 2:1. Pulse jet cleaning with air-to-cloth ratio of 6:1 costs 36% less at 10,000 acfm and 22% less at 100,000 acfm. Costs for pulse jet systems do not extend beyond 100,000 acfm.

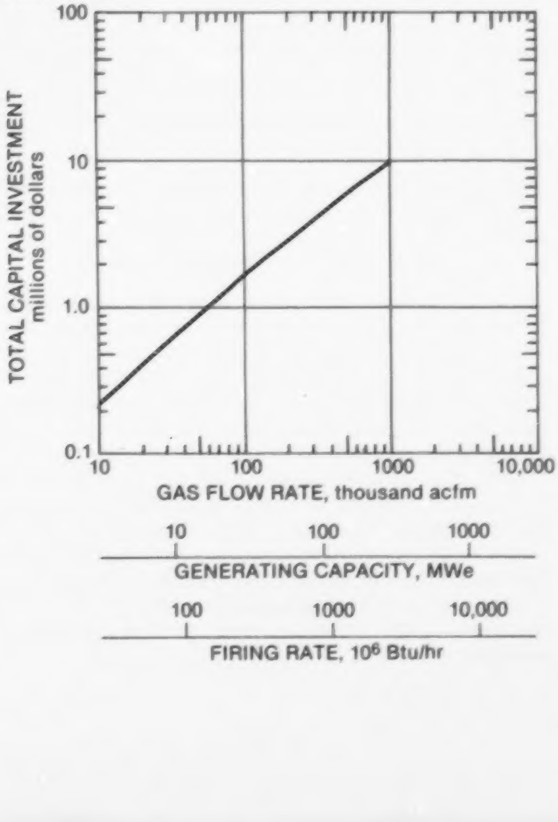
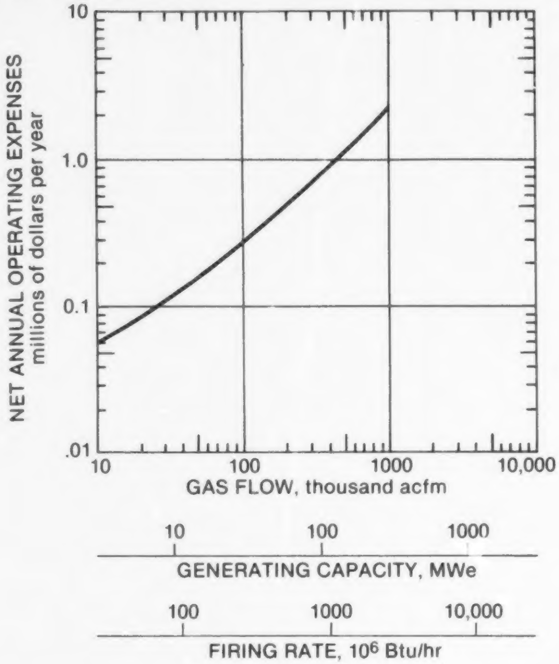


Figure 4-12. Fabric filter system for particulate matter collection - Net annual operating expenses (March, 1980 dollars).

Curves based on reverse air cleaning with air-to-cloth ratio of 2:1. Pulse jet cleaning with air-to-cloth ratio of 6:1 costs 9% less at 10,000 acfm and 14% less at 100,000 acfm. Costs for pulse jet systems do not extend beyond 100,000 acfm.



The net annual operating expense estimates are similarly based on costs in References 8, 10, 14, 16, and 18.

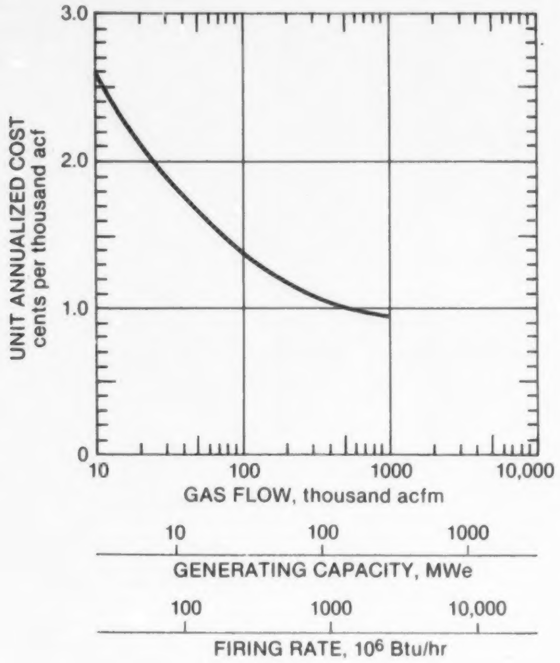
Major Variables Affecting Costs

Pulse jet baghouse systems are generally used in lower gas flow rate applications than reverse air baghouses. Since pulse jet cleaning operates at higher air-to-cloth ratios and smaller filter areas, it is less expensive than the other cleaning methods. Pulse jet systems require between 35 and 50 percent less capital investment than reverse air systems for applications treating 10,000 and 100,000 acfm. The net annual operating expenses for pulse jet systems are up to about 15 percent lower than for reverse air systems.

As discussed previously, the air-to-cloth ratio is determined from experience and can range from 1.5 to 3.5 fpm for reverse air cleaning and from 4 to 15 fpm for pulse jet cleaning. The case study fabric filter systems used to develop cost curves in this report are based on average or typical air-to-cloth ratios: 2 fpm for reverse air cleaning and 6 fpm for pulse jet cleaning. Systems featuring reverse air cleaning with

Figure 4-13. Fabric filter system for particulate matter collection - Unit annualized cost (March, 1980 dollars).

Curves based on reverse air cleaning with air-to-cloth ratio of 2:1. Pulse jet cleaning with air-to-cloth ratio of 6:1 is 17% less. Costs for pulse jet systems do not extend beyond 100,000 acfm.



an air-to-cloth ratio of 1 fpm (at a gas flow of 100,000 acfm) typically have total capital investment requirements about 45 percent higher than systems designed for an air-to-cloth ratio of 2 fpm, while net annual operating expenses are about 40 percent higher. Systems with air-to-cloth ratios of 4 fpm have total capital investment requirements about 25 percent lower than systems designed for an air-to-cloth ratio of 2 fpm, while net annual operating expenses are about 10 percent lower.

High inlet particle concentrations create the need for larger and more expensive ash disposal systems and increase the cost of ultimate disposal. Very high inlet particulate matter concentrations can also require increased filter areas or the use of a gas precleaning process. As an example, an inlet particulate matter concentration of 5 gr/acf or 2.5 times that specified for the model fabric filter systems would typically increase the capital investment requirement by about 5 percent but would increase net annual operating expenses by about 40 percent. On the other hand, an inlet particulate loading of 1 gr/acf would reduce the capital investment requirement by about 3 percent and reduce net annual operating expenses by about 15 percent. If the recovered dust could be recycled and disposal requirements minimized, annual operating expenses could be reduced further.

Expected filter bag life also affects costs. A reduction in bag life from 2 years (the value assumed in the cost curves) to 6 months would typically increase net annual operating expenses by up to about 20 percent. The impact of bag life is greatest on large systems where the costs of bag replacement are a relatively greater fraction of operating expenses.

Although the cost curves in Figures 4-11 through 4-13 are based on fabric filter systems applied to coal-fired boilers, they can be used for other applications if the inlet grain loadings, air-to-cloth ratios, and dust characteristics are similar. Fabric filtration generally can be used to collect dry particles from any source if moisture condensation can be avoided.

4.4 Venturi Wet Scrubbers

Description

A venturi wet scrubber is a collection device which uses an aqueous stream or slurry to remove particulate matter and/or gaseous pollutants.

Scrubbers are usually classified by energy consumption (in terms of gas-phase pressure drop). Low-energy scrubbers, represented by spray chambers and towers, have pressure drops of less than 5 in. H_2O. Medium-energy scrubbers, such as impingement scrubbers, have pressure drops of 5-15 in. H_2O. Higher-energy scrubbers, such as high-pressure drop venturi scrubbers, have pressure drops exceeding 15 in. H_2O. The most common scrubbers used for 'moderate' removals of particulate matter are medium-energy impingement and venturi scrubbers. Greater removals of particulate matter are usually achieved with high-energy venturi scrubbers.

The collection efficiency of scrubbers is essentially dependent on the characteristics of the particulate matter (particle size) and the energy input to the scrubber (as measured by pressure drop).

Venturi scrubbers have gained widespread popularity, especially for the collection of hygroscopic and corrosive submicron particles. In a typical venturi scrubber, the particle-laden gas first contacts the liquor stream in the core and throat of the venturi section. The gas and liquor streams then pass through the annular orifice formed by the core and throat, atomizing the liquor into droplets which are impacted by particles in the gas stream. Impaction results mainly from the high differential velocity between the gas stream and the atomized droplets. The particulate laden droplets then are removed from the gas stream by centrifugal action in a cyclone separator and, if appropriate, a mist elimination section.

Some major industries which use venturi scrubbers for particulate matter control are:

- Coal cleaning industry for emissions from coal handling systems.
- Phosphate fertilizer industry for emissions from all major sources except grinding and screening operations.
- Lime and asphalt plants for emissions from kilns.
- Metal (iron and steel, iron foundries, and ferroalloy) industries for emissions from various processing operations.

The equipment normally associated with venturi scrubber systems is shown in Figure 4-14.

As the particulate matter accumulates in the circulating scrubber liquid, a fraction of the liquid is removed and sent to disposal/reuse treatment. Fresh scrubber liquid is added to the circulating stream to replace evaporation losses and liquid removed as blowdown.

Because the pH of the scrubber circulating liquid may be altered due to the incidental absorption of acidic gas species such as SO_2, or collection of alkaline species such as lime dust, pH control of the blowdown liquor that is removed may be required before discharge.

Factors that affect the performance of typical wet scrubbers are:

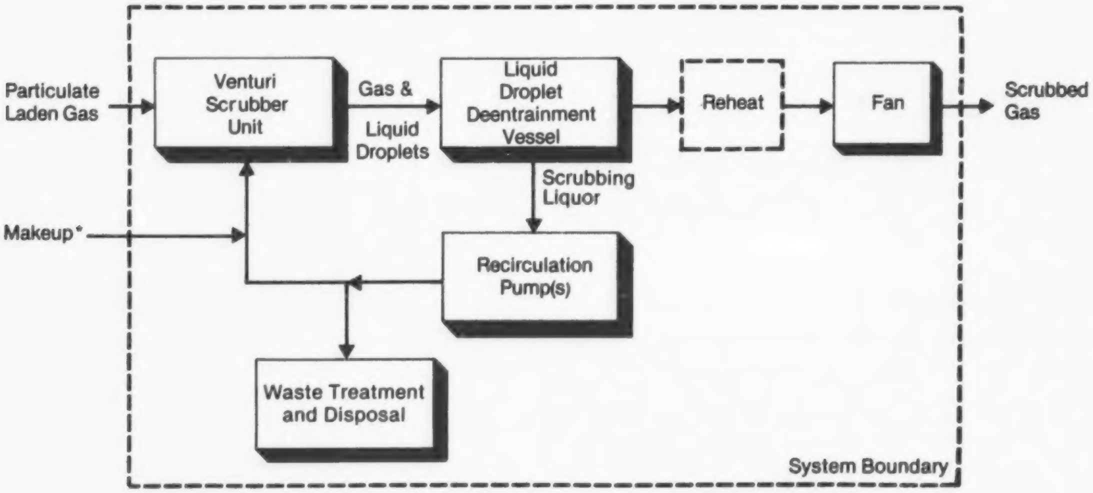
- Gas velocity (or gas phase pressure drop).
- Liquid-to-gas ratio (L/G).
- Particle size distribution.
- Inlet gas particulate matter concentration.

Although the performance of a scrubber depends directly on both the liquid-to-gas ratio and the gas velocity, the gas-phase pressure drop is usually the major factor affecting removal. As shown by Figure 4-15, removal efficiency increases with increasing gas-phase pressure drop: greater pressure drops (or gas velocities) create smaller liquid drops that are more efficient in collecting particulate matter. However, high-pressure drop scrubbers may show decreasing removal efficiency with increasing pressure drop due to carryover of the particulate-laden scrubbing droplets. High-pressure-drop scrubbers should thus be equipped with mist eliminators to ensure adequate separation of gas and scrubbing droplets.

If the liquid rate to the scrubber is sufficient to provide an adequate distribution of liquid droplets in the gas stream without flooding the scrubber, scrubber performance is relatively insensitive to variations in the liquid-to-gas ratio. Increases in the liquid-to-gas ratio generally increase scrubber efficiency, but the performance increases are usually small.

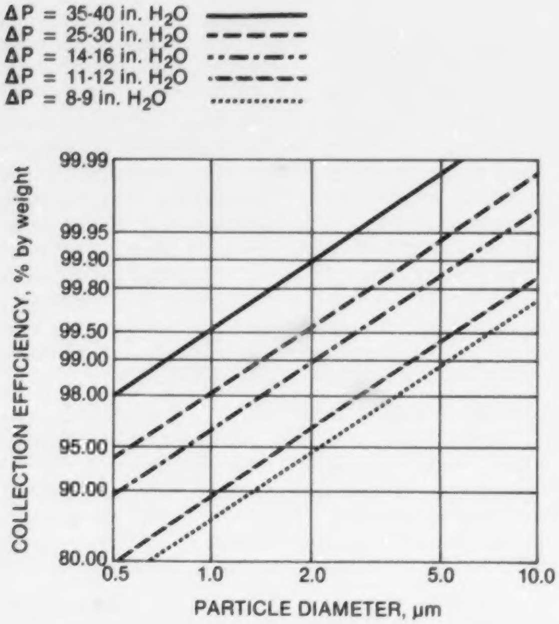
As shown in Figure 4-15, scrubber performance depends on the particle size distribution of the particle matter to be collected. This figure shows that collection efficiency varies directly with particle size, with larger particles collected at greater efficiency.

Figure 4-14. Venturi wet scrubber system for particulate matter collection.



*The amount of makeup added or liquor removed depends on the liquid losses due to evaporation into the gas stream, and the accumulation of dissolved solids in the recirculating liquid.

Figure 4-15. Venturi wet scrubber comparative fractional efficiency curves. (19)



Scrubber performance also depends on the particulate matter inlet concentration. Concentrations exceeding the scrubber design loading could overload the scrubber and reduce particulate matter removal efficiency. Scrubber efficiency could be improved by increasing the gas velocity (or pressure drop) and liquid-to-gas ratio. Alternatively, precleaners such as cyclones could be used upstream of the scrubber to reduce the particulate matter concentrations entering the scrubber.

Venturi scrubber applications generally include a variable throat system (enabling control of pressure drop) to enable a constant efficiency to be maintained at varying inlet conditions. For example, pressure drops across venturi throats generally range from 6 to 30 in. H_2O in boiler applications. Gas velocities through the venturi throat may range from 61 to 600 ft/s, while liquid-to-gas ratios vary from 8 to 15 gal/ 1000 ft^3 (8).

Design Basis and Costs

The design basis for costs presented in this summary includes the equipment items shown in Figure 4-14 and the design parameters of Table 4-8. Three scrubber pressure drops are assumed in order, to reflect differences in removal efficiency. The overall collection efficiency for a scrubber system will vary for specific particle sizes and properties.

Figures 4-16 and 4-17 present total capital investment and net annual operating expenses,

Table 4-8. Venturi Wet Scrubber Design Parameters

Parameters	Design basis	Typical range[a]
Gas flow rate, acfm	1,000 to 100,000	1,000 to 200,000
Inlet particulate matter concentration, gr/acf	2	Site specific
Pressure drop, in. H_2O	10, 20, and 40	6 to 80[b]
Operating pressure	Atmospheric	1 atmosphere and above
Liquid-to-gas ratio, gpm/acfm	20	10 - 20
Materials of construction	Carbon steel 316 Stainless steel	Carbon steel Fiberglass liners Stainless steel
Plate thickness, in.	1/8 to 5/16	Depends on scrubber size and pressure drop
Treatment and disposal	Clarification, solids hauling to landfill[c]	Site-specific
Slurry concentration to clarifier, % solids	15	Site-specific

[a] 200,000 acfm was largest scrubber capacity found in References 8, 10, and 14.

[b] Reference 20, p. 642.

[c] Design details for wastewater unit processes are available in Reference 21.

Figure 4-16. Venturi wet scrubber system for particulate collection - Total capital investment (March, 1980 dollars).

Costs based on carbon steel scrubber with clarifier for wastewater treatment and solids removal.

For stainless steel scrubber increase costs by about 134%.

A wastewater treatment system with vacuum filtration increases costs about 30% for a carbon steel scrubber system.

Costs between 1000 and 10,000 acfm are by graphical extrapolation.

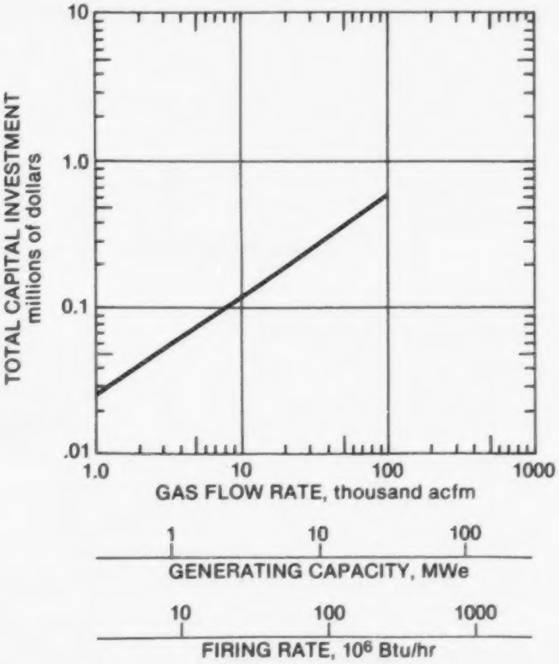


Figure 4-17. Venturi wet scrubber system for particulate collection - Net annual operating expenses (March, 1980 dollars).

Costs based on a carbon steel scrubber with clarifier for wastewater treatment and solids removal.

For stainless steel scrubber increase costs by about 24%.

A wastewater treatment system with vacuum filtration increases costs about 18%.

Costs between 1000 and 10,000 are by graphical extrapolation.

Pressure Drop (in. H_2O)	Energy	Collection Efficiency (2μm part.) %	
10	low	95	———
20	medium	99	– – – –
40	high	99.9	-·—·—·-

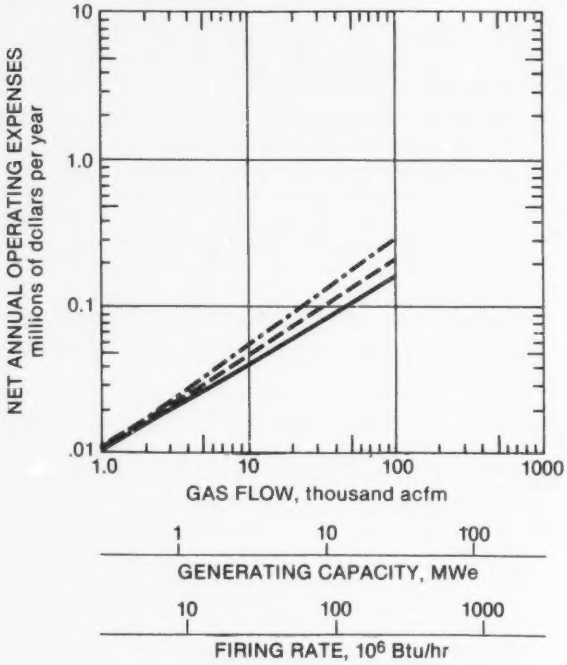


Figure 4-18. Venturi wet scrubber system for particulate collection - Unit annualized cost (March, 1980 dollars).

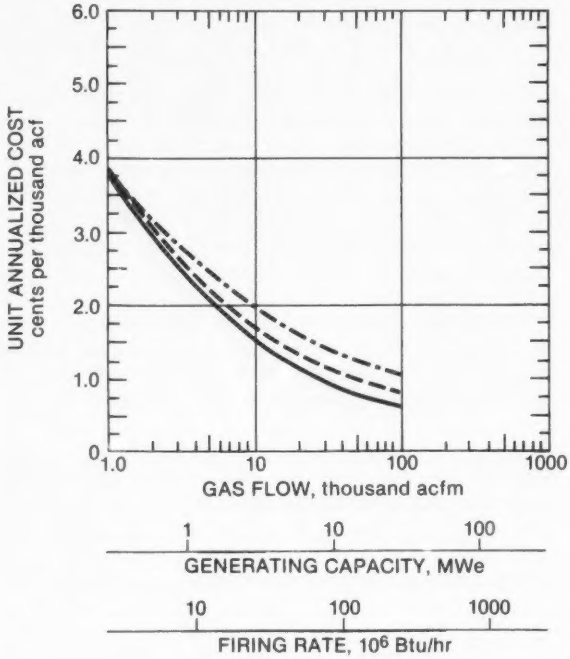
Costs based on carbon steel scrubber with clarifier for wastewater treatment and solids removal.

For stainless steel scrubber increase costs by about 52%.

A wastewater treatment system with vacuum filtration increases costs about 34% for a carbon steel scrubber system.

Costs between 1000 and 10,000 acfm are by graphical extrapolation.

Pressure Drop (in. H_2O)	Energy	Collection Efficiency (2μm part.) %	
10	low	95	———
20	medium	99	– – – –
40	high	99.9	-·—·—·-



respectively, for wet scrubber systems including costs for wastewater treatment and solids disposal. Figure 4-18 presents wet scrubber system unit annualized costs.

A number of possible configurations can be used for wastewater treatment depending on solids' settling characteristics and volume of wastewater treated.

In some scrubber systems a simple settling tank (sometimes integral with the physical structure of the scrubber) is sufficient. Solids settle easily and are removed from the bottom of the vessel as a sludge which is hauled "as is" to disposal. Clarified water from the tank is recycled to the scrubber. Some fractional discharge of the total scrubber water flow will be required, however, due to a gradual buildup of dissolved solids in the scrubbing solution. This may or may not require discharge treatment depending on the effluent regulations that apply to the specific site.

Other scrubber systems might employ a separate clarifier-thickener with rotating internals to enhance solids settling and scour sludge solids from the vessel bottom into the vessel sludge discharge port. The clarification thickening process might employ a chemical additive system. Certain polymers can be used to enhance solids settling. Thickened sludge would then be pumped to storage for subsequent hauling to landfill. Or, the sludge might first be vacuum filtered to reduce the water content prior to ultimate disposal. It is clear that the total costs for a wet scrubber system might vary significantly depending on the mode of wastewater treatment. In this report, the typical system uses a clarifier for treatment. The costs figures provide factors for relative costs of both simpler and more complex scrubber wastewater treatment systems.

Major Variables Affecting Costs

A major variable affecting costs is the operating pressure drop. Its effects on costs are clearly shown in Figures 4-17 and 4-18. Other variables include equipment design details.

The capital cost of the scrubber and de-entrainment vessel is based on the volumetric flow rate, operating pressure, and construction materials. The sizes of the scrubber unit and de-entrainment vessel are determined from the actual inlet gas flow rate. Additional factors determining capital costs are metal thicknesses (which depend on erosion, corrosion, and pressure drop), mode of scrubber pressure drop control (whether by manual or automatic venturi throat), and construction materials (stainless steel, fiberglass, or rubber liners). The plate thickness for the scrubber and separator is a function of the maximum operating design pressure and vessel diameter. As the volume flow rate and/or pressure drop increase, the metal wall thicknesses must also be increased to prevent buckling. Some allowances for corrosion or erosion are usually added to the design specifications. Additional plate thickness becomes necessary at higher air flow rates.

Pressure drop has a significant effect on operating expenses because it determines power costs for gas flow. The pressure drop through the venturi is a function of gas stream throat velocity and to a lesser extent the scrubbing liquor flow rate. The desired collector efficiency determines the magnitude of the above variables with smaller particle sizes, requiring higher pressure drops. The pressure drop can be increased by increasing the gas stream throat velocity, increasing the scrubbing liquor flow rate, or both. However, the effectiveness of increasing the liquid flow rate diminishes significantly after a point.

Another very significant cost factor is liquid and solid waste disposal. Exact requirements depend on the specific characteristics of the solids collected and the corrosivity of the liquid stream. Not only can the method of wastewater treatment vary, but there are many options for ultimate solids disposal. For the scrubber systems addressed in this report, wastewater treatment and solids disposal costs are about .5 percent to 30 percent of total capital investment and 9 percent to 30 percent of net annual operating expenses for systems treating 10,000 acfm to 100,000 acfm, respectively, with a 2 gr/acf inlet grain loading.

The inlet gas particulate concentration also impacts venturi scrubber costs. Although inlet concentration is directly related to pressure drop, there are limits to venturi scrubber applications. Very high particulate loadings (e.g., 15 gr/dscf for lime kilns) may be more cost-effectively controlled if a mechanical collector (cyclone) is used upstream of the venturi scrubber. This collector reduces the particulate loading to the scrubber, thereby reducing scrubber erosion and pressure drop (reduced capital/operation costs).

Venturi scrubbers are not as efficient as fabric filters and ESP's for submicron particle collection. Therefore, gas streams with loadings or high submicron particle loadings may be more effectively controlled with fabric filters or ESP's.

References - Section 4

1. McIlvaine Company (Northbrook Illinois). The Electrostatic Precipitator Manual, Vol. I. 1976.
2. U.S. Environmental Protection Agency. Appendix A: Compilation of Air Pollutant Emission Factors, AP-42. Office of Air Quality Planning and Standards, Research Triangle Park, NC. February 1980. p. A-2.
3. U.S. Department of Commerce, Bureau of the Census. Current Industrial Reports, Pollution Abatement Costs and Expenditures, 1978, MA-200(78-2). p. 24.
4. Sittig, M. Particulates and Fine Dust Removal, Processes and Equipment. Park Ridge, NJ, NOYES Data Corporation, 1977. p. 10.
5. Buonicore, Anthony J. Air Pollution Control, Chemical Engineering *87*(13) p. 83.
6. Theodore, L., and A.J. Buonicore. Industrial Air Pollution Control Equipment for Particulates. CRC Press: Cleveland, OH, 1976.
7. Webber, D. (Joy/Western Precipitation Division). Telephone communication with P.J. Murin (Radian Corporation). September 11, 1980.
8. Neveril, R.B. (GARD, Inc.) Capital and Operating Costs of Selected Air Pollution Control Systems. EPA-450/5-80-002, U.S. Environmental Protection Agency, December 1978.
9. Tighe, S.C. Mechanical Collectors for Particulate Control, Stoker Coal-Fired Boilers (Radian Corporation Internal Technical Note). October 3, 1980.
10. PEDCo Environmental, Inc. Capital and Operating Costs of Particulate Controls on Coal- and Oil-Fired Industrial Boilers. EPA-450/5-80-009, U.S. Environmental Protection Agency, Research Triangle Park, NC, August 1980.
11. Oglesby, S., Jr. and G.B. Nichols. (Southern Research Institute) A Manual of Electrostatic

Precipitator Technology, Part I — Fundamentals. EPA No. APTD 0610, PB 196 380*, National Air Pollution Control Administration, Cincinnati, OH, August 1970.

12. Oglesby, S., Jr. and G.B. Nichols. (Southern Research Institute) A Manual of Electrostatic Precipitator Technology, Part II — Application Areas. EPA No. APTD 0611, PB 196 381*, National Air Pollution Control Administration, Cincinnati, OH, August 1970.
13. Danielson, John A. (Ed.) Air Pollution Engineering Manual, 2nd Ed. EPA No. AP-40, PB 225 132*, Los Angeles County Air Pollution Control District, Los Angeles, CA. May 1973.
14. Industrial Gas Cleaning Institute. Air Pollution Control Technology and Costs in Seven Selected Areas. EPA-450/3-73-010, PB 231 757*, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, December 1973.
15. Loudin, D.J. (Industrial Gas Cleaning Institute, Inc.) Electrostatic Precipitator Costs for Large Coal-Fired Steam Generators. EPA-450/3-78-045, PB 290 169*, U.S. Environmental Protection Agency, Research Triangle Park, NC, February 1977.
16. Campbell, K.S., et al. (Stearns-Roger Corporation) Economic Evaluation of Fabric Filtration Versus Electrostatic Precipitation for Ultrahigh Particulate Collection Efficiency. EPRI-FP-775, EPRI RP 834-1, Electric Power Research Institute, Palo Alto, CA, June 1978.
17. Billings, C.E. and J. Wilder (GCA Corporation) Handbook of Fabric Filter Technology, Vol I. Fabric Filter Systems Study. EPA No. APTD-0690, PB 200 648*, U.S. Department of Health, Education and Welfare, Washington, DC, December 1970.
18. Evans, R.J. Methods and Costs of Dust Control in Stone Crushing Operations. IC-8669, U.S. Bureau of Mines, Pittsburgh, PA, January 1975.
19. Joy Industrial Equipment Company. Western Precipitation Gas Scrubbers: Type "V" Turbulaire Variable Venturi Scrubber. Los Angeles, CA, Joy Manufacturing Company, 1978. 6 pp.
20. Liptak, B.G., (ed.), Environmental Engineer's Handbook, Air Pollution, Vol. 2. Chilton Book Company, Radnor, PA, 1974.
21. Office of Water Program Operations. Innovative and Alternative Technology Assessment Manual. EPA-430/9-78-009, U.S. Environmental Protection Agency, Cincinnati, OH, February 1980.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

Section 5
Flue Gas Desulfurization

Sulfur oxide (SO_x) emissions (primarily sulfur dioxide, SO_2) arise from fuel combustion and some industrial processes. Nationwide emissions of SO_x were about 30 million tons in 1978 (1).

Sources of SO_2 emissions are regulated by both federal and state governments. Federal new source performance standards are now in effect for new fossil-fuel-fired electric power plants. Existing emission sources are covered by state implementation plans mandated under the Clean Air Act.

Flue gas desulfurization (FGD) scrubbing systems are used to absorb SO_2 gas from combustion gases to meet emission regulations. This section presents a summary of total capital investment, net annual operating expenses, and unit annualized costs for FGD systems of the type applied to utility boiler and industrial boiler stack gases. These systems can also be used to reduce emissions from other industrial processes such as smelter operations and process furnaces.

Two general categories of FGD systems are wet scrubbing and spray drying systems. The features of these two categories are discussed separately in the following pages.

5.1 Wet FGD Scrubbing Processes

The wet FGD processes presented in this report have been selected because they are the most extensively used processes and are expected to continue to be used for future installations.

Wet FGD processes considered here include:
- Lime/limestone.
- Non-regenerable sodium alkali scrubbing (throw-away).
- Dual alkali.
- Magnesium oxide.
- Wellman-Lord.

The first three are non-regenerable processes. They produce a major waste stream that requires disposal. The lime/limestone and dual alkali systems produce a solid waste stream; the non-regenerable sodium alkali scrubbing system produces a liquid waste stream.

The magnesium oxide and Wellman-Lord systems are regenerable processes which produce a saleable product instead of a major waste stream. Both processes yield a concentrated SO_2 stream which can be used for liquefied SO_2, sulfuric acid, or elemental sulfur production.

Costs for utility boiler applications are presented for the lime/limestone, dual alkali, magnesium oxide, and Wellman-Lord processes. Costs for non-regenerable sodium alkali scrubbing applied to utility boilers are not included as this process is used primarily on smaller industrial boilers. Costs for FGD systems on industrial boilers are presented for the limestone, dual alkali, Wellman-Lord, and non-regenerable sodium alkali scrubbing processes.

A brief process description is included for each of these wet FGD processes in Section 5.1.1. Section 5.1.2 discusses the design bases and costs derived from the major cost references used for this summary. The cost curves presented show total capital investment, net annual operating expenses, and unit annualized costs. Section 5.1.3 provides a discussion of some of the variables which have a major impact on costs.

5.1.1 Wet FGD Process Descriptions

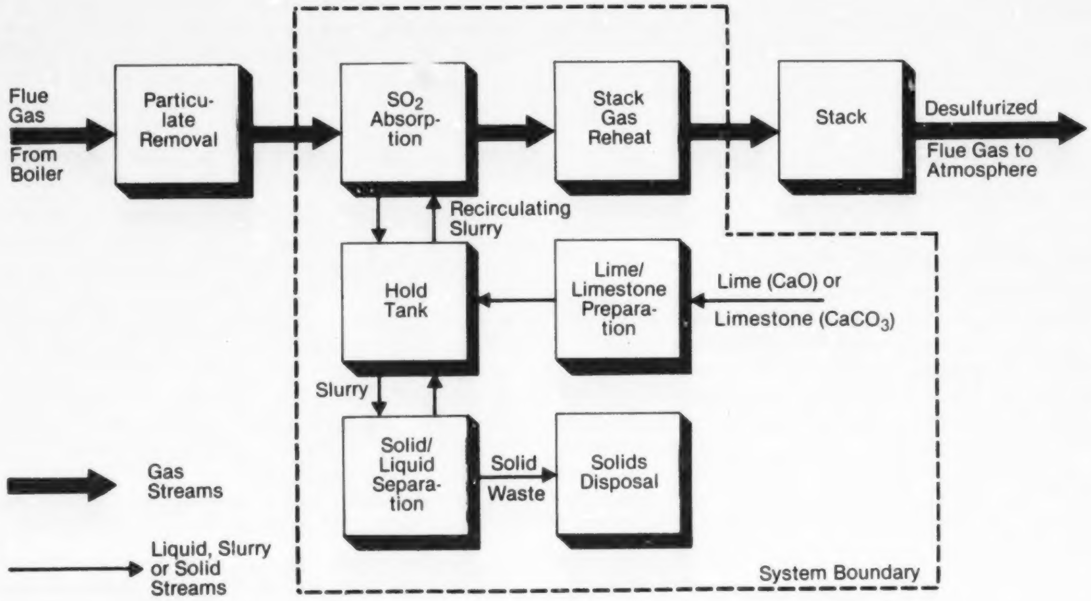
This section contains a brief process description for each of the FGD systems contained in this summary. Stack gas reheat (SGR) is shown in each FGD system. SGR has been used following wet FGD processes to protect downstream equipment against corrosion and to achieve better plume dispersion.

Lime/Limestone Scrubbing Process

In the lime/limestone process, a slurry containing calcium hydroxide or calcium carbonate removes SO_2 from flue gas in a wet scrubber. Both of these processes are non-regenerable processes and produce a large volume of solid waste for disposal. The lime and limestone systems are considered together here because of their similarity. A block diagram showing the major processing areas in lime/limestone scrubbing is given in Figure 5-1.

Particulate matter is normally removed upstream of the FGD system. After particulate matter removal, the flue gas enters the scrubber where SO_2 is absorbed by contact with a slurry of lime or limestone in water. The SO_2 chemically reacts with the lime or limestone to form calcium sulfite and calcium sulfate.

Figure 5-1. Lime/limestone scrubbing process for flue gas desulfurization.


Flue Gas
From Boiler
Particulate Removal
SO2 Absorption
Stack Gas Reheat
Stack
Desulfurized
Flue Gas to Atmosphere
Recirculating Slurry
Hold Tank
Lime/ Limestone Preparation
Lime (CaO) or
Limestone (CaCO3)
Slurry
Solid/ Liquid Separation
Solid
Waste
Solids Disposal
Gas Streams
Liquid, Slurry or Solid Streams
System Boundary

Calcium sulfite and sulfate crystals are only slightly soluble in water and precipitate from solution. The slurry passes through the scrubber into a hold tank. The hold tank is designed to allow enough time for solids precipitation to proceed. Fresh lime or limestone is also added to this tank. Most of the slurry is recirculated to the scrubber. The remainder is continually removed from the hold tank for solid/liquid separation by ponding or clarification/vacuum filtration processes. In ponding, the pond serves as the final solids disposal method as well as liquid clarifier. Solids from filtration are usually disposed of in a landfill. In either case, the clarified liquid is returned to the scrubber system for reuse.

The lime/limestone process is by far the most extensively used FGD system for utility boilers, representing more than 80 percent of FGD units in operation or under construction (2). By contrast, less than 5 percent of the industrial boiler FGD installations are lime/limestone (3). The non-regenerable sodium alkali scrubbing process, described in the next section, has been preferred in small boiler applications partly because of its simplicity, low maintenance requirements, and low total capital investment.

Non-regenerable Sodium Alkali Scrubbing (Throwaway) Process

In sodium alkali scrubbing systems, a scrubbing solution of sodium hydroxide, sodium carbonate, or sodium bicarbonate absorbs SO_2 from flue gas. A block diagram showing the various sodium scrubbing process modules is presented in Figure 5-2.

The SO_2 chemically reacts to form sodium sulfite and sodium bisulfite which remain dissolved in solution. Part of the sulfite in solution reacts with oxygen from the flue gas to form sodium sulfate. The sodium sulfite and sulfate salts are removed from the system in solution as a liquid waste. Sodium carbonate (Na_2CO_3) or sodium hydroxide (NaOH) are generally selected as the makeup sodium alkali added to the recirculating scrubber solution to compensate for the quantity that reacts with SO_2.

Sodium alkali scrubbing differs from limestone and dual alkali FGD systems in that no solid waste product is formed. However, larger quantities of liquid waste containing sodium sulfite, sodium bisulfite, and sodium sulfate must be disposed of. Disposal practices for this waste stream include wastewater treatment, holding ponds for evaporation, and deep-well injection.

The non-regenerable sodium alkali scrubbing process is the simplest FGD process described in this report from the standpoint of operation and maintenance. Even so, only four utility systems are operational (4). This is largely due to the liquid waste disposal problem and the high cost of the sodium chemicals required for scrubbing. However, for industrial boilers, non-regenerable sodium alkali scrubbing has been the preferred process to date with over 100 operational systems (3). Many of these systems are very small (treating flue gas from a boiler with a firing rate of less than 100 x 10^6 Btu/hr). In these systems, a premium has been placed on simplicity, low capital cost, and reliability during operation. The non-regenerable sodium alkali scrubbing process meets these requirements and has found extensive use in spite of high chemical costs. However, future application of this process is limited in many areas of the country due to the high cost of treating the liquid waste for final disposal.

Dual Alkali Scrubbing Process

The dual alkali process encompasses some features of both the non-regenerable sodium alkali scrubbing and lime FGD processes. The scrubbing liquid is a solution of soluble sodium-based alkali containing sodium carbonate, sodium bicarbonate, sodium sulfite, and sodium hydroxide. Calcium-based solids, similar to those formed in lime/limestone systems, are produced by addition of lime to a stream of spent scrubbing liquor. A block diagram showing the various dual alkali process modules is presented in Figure 5-3.

After removal of particulate matter, the flue gas enters the scrubber where SO_2 is absorbed. The SO_2 reacts chemically with sodium alkali to form sodium sulfite and sodium bisulfite. Some of the sodium sulfite reacts in solution with oxygen (O_2) from the flue gas to form sodium sulfate.

A side stream of scrubbing solution is reacted with slaked lime to form calcium sulfite and calcium sulfate which precipitate from solution. This process step also results in regeneration of the sodium-based alkali for recycle to the scrubber. The regenerated solution contains sodium sulfite and sodium hydroxide.

The side stream slurry, which contains calcium sulfite and sulfate solids, is sent to a thickener where solids are concentrated by sedimentation. This forms a sludge of solids and water which is further thickened in a vacuum filter and then washed to recover sodium salts. The solids are then disposed of by either ponding or landfill. Clarified solution from these steps is returned to the scrubber.

Dual alkali systems can be classified as either dilute mode or concentrated mode processes, depending on the concentration of alkali in the scrubbing liquid. The choice between these two operations is determined by how much sulfite reacts with oxygen as discussed above. The quantity of sulfite that reacts depends on site-specific factors such as the relative concentrations of SO_2 and O_2 in the flue gas. The sodium sulfate formed must be removed from the system to maintain SO_2 removal capability, and precipitation of sodium sulfate in the clear liquid circulation loop. The two operations differ in the manner in which sulfate is removed from the system.

The dilute mode system is applicable in systems where high oxidation rates are expected. Firing of low sulfur coal produces flue gas characteristics for which the dilute mode is better suited. Concentrated mode systems are applicable where high sulfur coal is encountered. Most existing dual alkali systems on both utility boilers and industrial boilers use the concentrated mode (3).

Magnesium Oxide Scrubbing Process

The magnesium oxide or magnesia slurry absorption process is a regenerable process which uses

Figure 5-2. Sodium alkali scrubbing (throwaway) process for flue gas desulfurization.

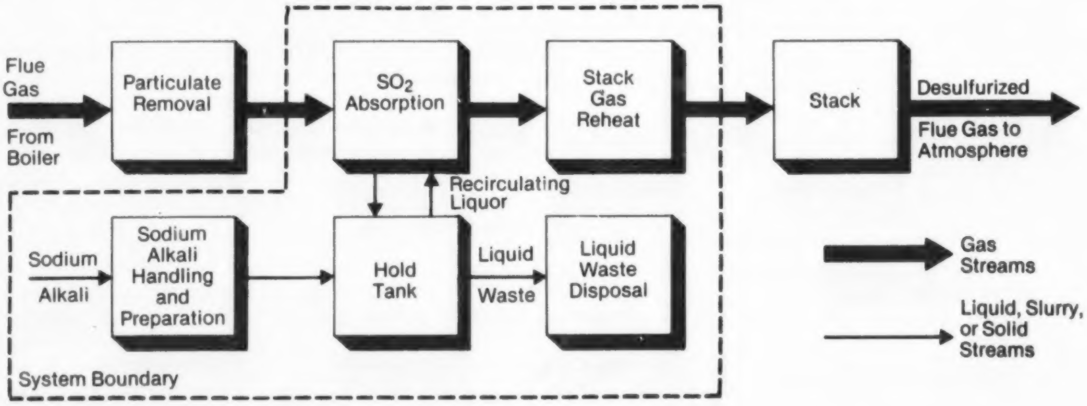
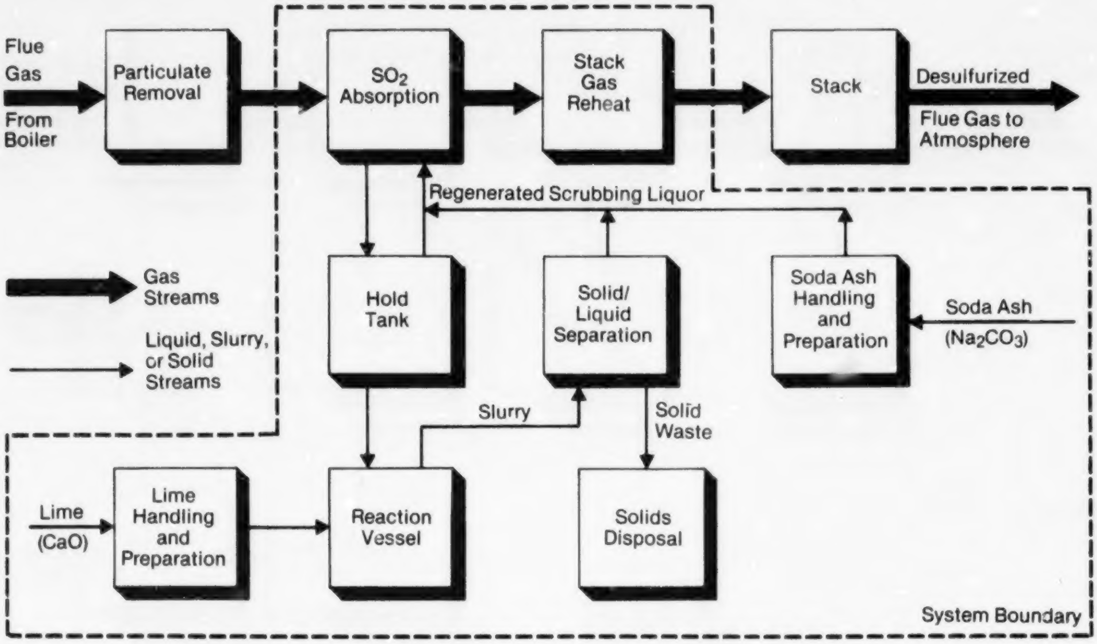


Figure 5-3. Dual alkali scrubbing process for flue gas desulfurization.


Flue Gas From Boiler
Particulate Removal
SO2 Absorption
Stack Gas Reheat
Stack
Desulfurized Flue Gas to Atmosphere
Regenerated Scrubbing Liquor
Gas Streams
Liquid, Slurry, or Solid Streams
Hold Tank
Solid/ Liquid Separation
Soda Ash Handling and Preparation
Soda Ash (Na2CO3)
Slurry
Solid Waste
Lime (CaO)
Lime Handling and Preparation
Reaction Vessel
Solids Disposal
System Boundary

magnesium hydroxide to absorb SO_2 in a wet scrubber. A block diagram showing the major processing areas for the magnesium oxide process is presented in Figure 5-4.

After particulate removal, the flue gas may be pretreated in a pre-scrubber for chloride removal to prevent excess concentrations of chloride in the scrubbing liquid. Then the flue gas is contacted with magnesia slurry in a scrubber to remove SO_2. The SO_2 reacts with the magnesium hydroxide to form magnesium sulfite and magnesium bisulfite. Some of the sulfite formed in the above reactions reacts with oxygen from flue gas to form magnesium sulfate.

Spent slurry from the scrubber is sent to a hold tank designed to provide sufficient holding time for the solids to precipitate. The sulfite and sulfate solids precipitate as hydrated crystals.

The hold-tank effluent is split into two streams. The first stream of relatively clear liquid is combined with fresh magnesium oxide and recycled slurry. This stream is then recycled to the scrubber.

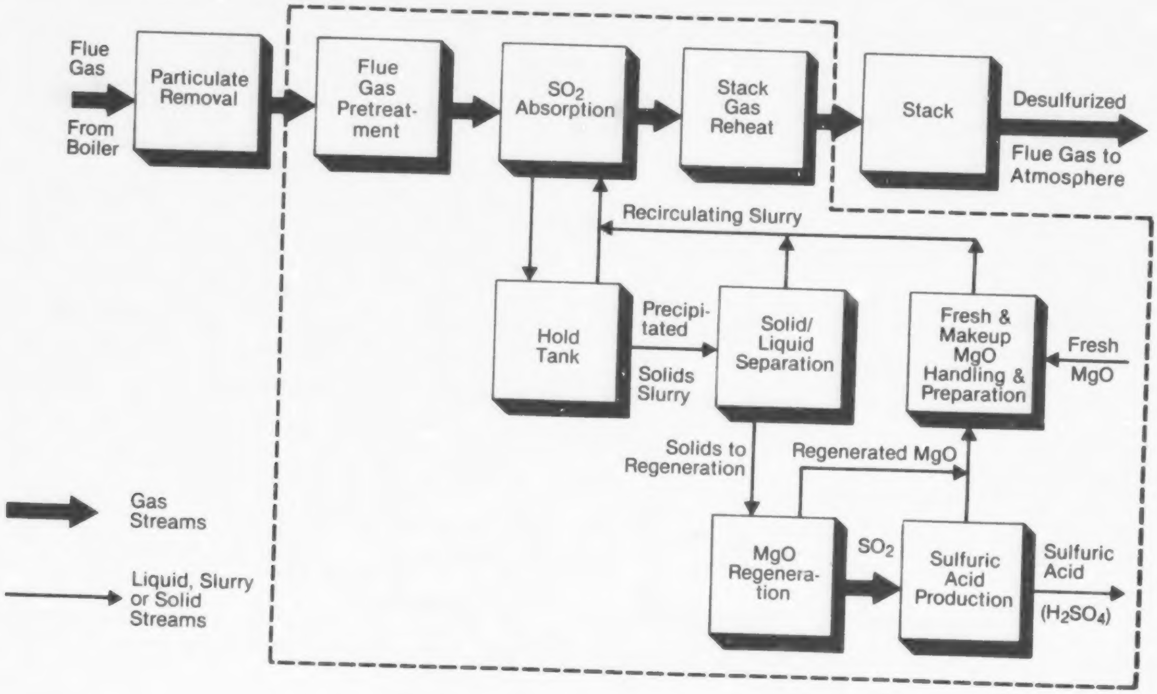
The second stream of liquid and settled solids is sent to a thickener for solids concentrations. After further solids concentration in a centrifuge, the magnesium sulfite and magnesium sulfate hydrated crystals are regenerated to produce fresh magnesium oxide. The regeneration step takes place in two stages. First, water is driven off in an oil-fired rotary kiln. Then the dried magnesium sulfite and magnesium sulfate crystals are calcined. During calcining, magnesium sulfite and magnesium sulfate are converted to magnesium oxide with the evolution of SO_2 gas. The off-gas from the calciner contains from 8 to 10 percent SO_2 and is used in a downstream recovery process such as sulfuric acid production (3).

The magnesium oxide process has been shown to be feasible on a full-scale utility boiler system. Three units, all retrofits in the 95 to 150 MWe size range, have demonstrated greater than 90 percent SO_2 removal on both coal-fired and oil-fired systems. At present, however, only one of these units is in operation (4). There are no applications of the magnesium oxide system on industrial boilers in the U.S. (3).

Wellman-Lord Scrubbing Process

The Wellman-Lord scrubbing system is a regenerable process that uses a sodium sulfite solution to absorb SO_2. It produces a concentrated stream of SO_2 that can be processed into elemental sulfur, sulfuric acid, or liquid SO_2. A block diagram showing the major

Figure 5-4. Magnesium oxide scrubbing process for flue gas desulfurization.


Flue Gas
From Boiler
Particulate Removal
Flue Gas Pretreat-ment
SO2 Absorption
Stack Gas Reheat
Stack
Desulfurized
Flue Gas to Atmosphere
Recirculating Slurry
Hold Tank
Precipi-tated
Solids Slurry
Solid/ Liquid Separation
Fresh & Makeup MgO Handling & Preparation
Fresh
MgO
Solids to Regeneration
Regenerated MgO
MgO Regenera-tion
SO2
Sulfuric Acid Production
Sulfuric Acid
(H2SO4)
Gas Streams
Liquid, Slurry or Solid Streams

processing areas required for Wellman-Lord scrubbing is presented in Figure 5-5.

In addition to the usual removal of particulate matter, the flue gas is also pretreated in a venturi scrubber for chloride removal to prevent excessive chloride concentrations in the scrubbing liquid.

Humidified gas from the prescrubber enters the absorption tower where it is contacted with the scrubbing liquor. Sodium sulfite reacts in solution with oxygen from the air to form sodium sulfate. The sulfate must be removed from solution in order for the scrubbing liquid to maintain its ability to absorb SO_2. A portion of the spent scrubbing liquid is sent to a treatment step which may employ either a heated sulfate crystallizer or a refrigerated chiller-crystallizer. Both produce a slurry of sodium sulfate solid crystals which is centrifuged. The resulting cake of solids is dried with steam, and disposed of as solid waste.

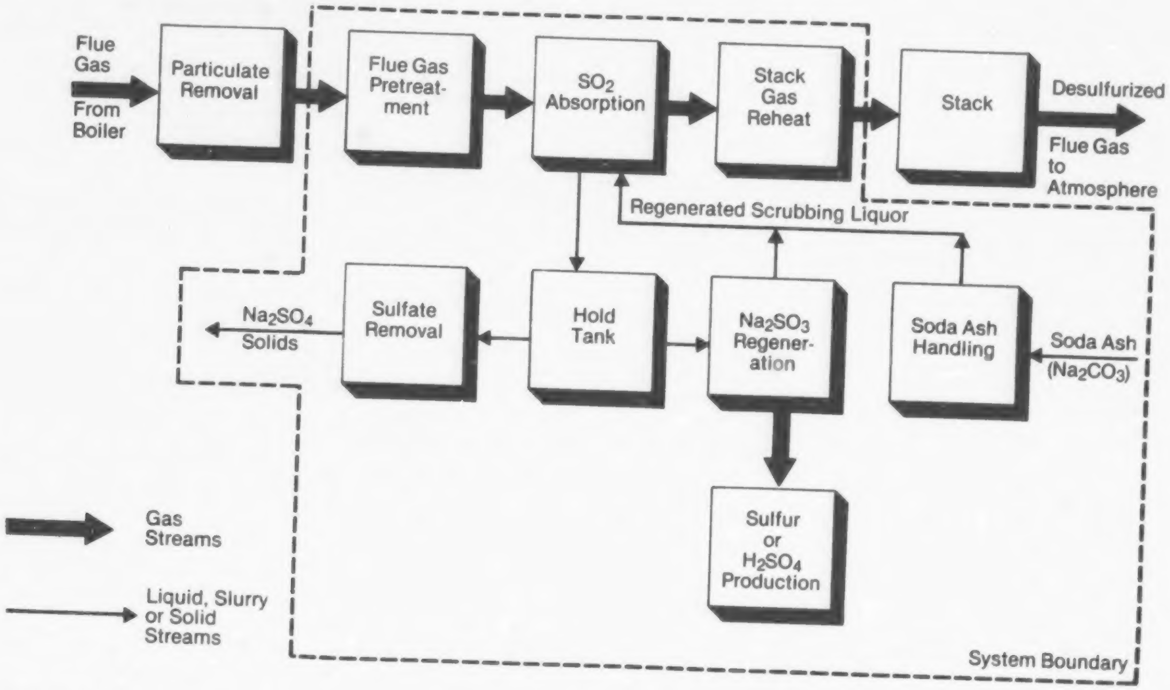
The remainder of the spent scrubbing liquid is regenerated by converting sodium bisulfate to sodium sulfite. This is accomplished by heating the liquid in a set of double-effect forced-circulation evaporators. Sodium sulfite crystals formed during this regeneration step are dissolved in water and recycled to the SO_2 absorber. The concentrated SO_2 stream may be used for production of elemental sulfur, sulfuric acid, or liquid SO_2.

Currently, there are seven Wellman-Lord systems installed on utility boilers in the U.S. Three of these systems produce elemental sulfur as a by-product, while the other four produce sulfuric acid (4). The predominant use of the Wellman-Lord system in the manufacturing industrial sector is for removal of SO_2 from Claus plant and sulfuric acid plant tail gas. These systems desulfurize streams having SO_2 concentrations ranging from 2700 to 10,000 ppm, higher than that normally found in boiler flue gas. The high alkalinity of the Wellman-Lord scrubbing solution achieves good removal of SO_2 from these concentrated gas streams. Currently there are no Wellman-Lord systems installed on industrial coal-fired boilers (3).

5.1.2 Design Basis and Costs

Costs of wet FGD systems for utility boilers and industrial boilers are presented in this section. Cost data for utility boiler applications were available from two primary sources: estimates developed by the Tennessee Valley Authority (2, 5), and estimates by PEDCo Environmental, Inc. (6). Costs adapted from the latter source only have been used as the bases of cost curves presented in this report, as explained shortly. Costs for industrial boiler applications were obtained from estimates developed by Radian Corporation (3). These sources were selected, in part, because each contains cost estimates for most of the

Figure 5-5. Wellman-Lord scrubbing process for flue gas desulfurization.


Flue Gas
From Boiler
Particulate Removal
Flue Gas Pretreatment
SO_2 Absorption
Stack Gas Reheat
Stack
Desulfurized
Flue Gas to Atmosphere
Regenerated Scrubbing Liquor
Na_2SO_4 Solids
Sulfate Removal
Hold Tank
Na_2SO_3 Regeneration
Soda Ash Handling
Soda Ash (Na_2CO_3)
Sulfur or H_2SO_4 Production
Gas Streams
Liquid, Slurry or Solid Streams
System Boundary

types of FGD systems evaluated in this study and because the costs were well documented. In addition, each of these references is relatively recent and each contains information on the effect of important variables on capital investment and annual operating expenses.

The design bases used in these references differ in some important respects. Some of the more significant factors are listed in Table 5-1. The design bases (and resultant costs) published by TVA reflect early FGD design concepts that may not apply to future installations. They do however, provide a baseline for illustrating relative cost effects due to differences in design basis as discussed in Section 5.1.3.

Cost for wet FGD systems are presented in Figures 5-6 through 5-11 for utility and industrial boilers. Capital costs are expressed as total capital investment. The curves reflect the cost of a new FGD system treating flue gas from a boiler burning 3.5 percent sulfur coal. Included in the total capital investment are costs for the purchase and installation of all equipment in each of the process areas within the system boundary defined in Figures 5-1 through 5-5 and described in Table 5-1. Annual operating costs are expressed as net annual operating expenses and

Table 5-1. Comparison of Design Bases for Major Cost References

TVA (2, 5) Utility boiler application	PEDCo (6) Utility boiler application	Radian (3) Industrial boiler application
	Design factors common to all FGD systems	
● Cost estimates are for new units. ● Size range from 200 to 1000 MWe (1840×10^6 to 8700×10^6 Btu/hr). ● Coal contains 3.5 percent sulfur and has a heating value of 10,500 Btu/lb. ● System designed for allowable emissions of 1.2 lb/10^6 Btu (78.5 percent removal for a 3.5 percent sulfur coal).	● Cost estimates are for new units. ● Size range from 100 to 1000 MWe (950×10^6 to 8700×10^6 Btu/hr). ● Coal contains 3.5 percent sulfur and has a heating value of 12,000 Btu/lb. ● System designed for 90 percent removal of maximum anticipated 3-hour average coal sulfur content.	● Cost estimates are for new units. ● Size range from 30×10^6 to 400×10^6 Btu/hr. ● Coal contains 3.5 percent sulfur and has a heating value of 11,800 Btu/lb. ● System designed for 90 percent removal.

(Continued)

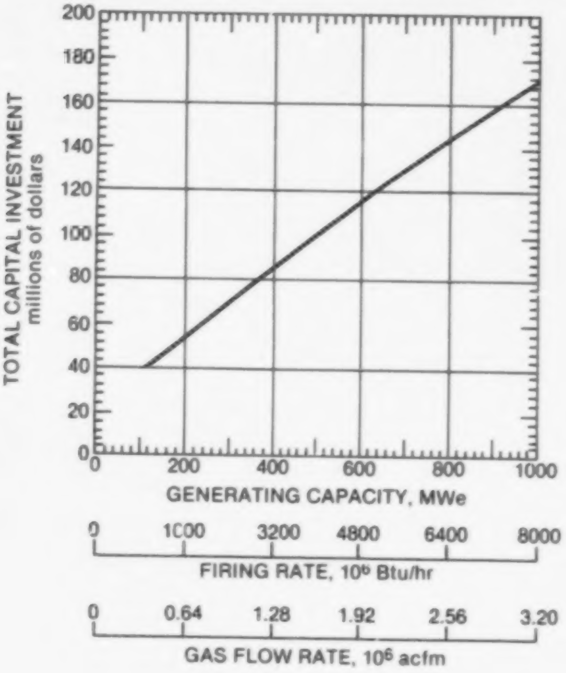
Table 5-1 (Continued)

TVA (2, 5) Utility boiler application	PEDCo (6) Utility boiler application	Radian (3) Industrial boiler application
● No process redundancy except for spare pumps.	● One spare scrubber, excess in-process storage capacity, and spare pumps.	● No process redundancy except for spare pumps.
● Stack gas reheat to 175°F.	● Stack gas reheat to 175°F.	● No stack gas reheat.
● Operating capacity factor is 0.80.	● Operating capacity factor is 0.65.	● Operating capacity factor is 0.60.
● Heat rate, Btu/kWhe-9200 at 200 MWe, 9000 at 500 MWe, and 8700 at 1000 MWe.	● Heat rate, Btu/kWhe - 9500 at 100 MWe, 9200 at 200 MWe, 9000 at 500 MWe, and 8700 at 1000 MWe.	● Not applicable.
● ESP and associated induced draft fans not included in cost. Cost of forced draft fan (relative to FGD unit) included for each scrubber train.	● Particulate matter removal costs not included.	● Particulate matter removal costs not included. Cost of fan for FGD system is included.
● Costs include all ductwork associated with FGD unit.	● Costs include all ductwork associated with FGD unit.	● Costs include all ductwork associated with FGD unit.
● No gas bypass provisions.	● Complete gas bypass provisions.	● No gas bypass provisions.
● 30-year life.	● 35-year life.	● 15-year life.
● Midwest location.	● Midwest location.	● Midwest location.
Design factors specific to lime/limestone systems		
● Scrubbers are turbulent contact absorber (TCA) variety.	● Scrubbers are turbulent contact absorber (TCA) variety.	● Scrubber is turbulent contact absorber (TCA) variety.
● Solids disposal method is ponding of hold tank slurry. Capital costs include cost of pond construction and 1-mile pipeline for transport of slurry to pond.	● Solids disposal method is ponding of thickened slurry after stabilization with lime and fly ash. Capital costs include clarifiers and on-site pond construction.	● Solids disposal method is off-site landfill. Disposal cost is $15.00/ton at 50 percent solids. Capital costs include costs for clarifiers and filters but not for landfill site preparation.
Design factors specific to sodium alkali scrubbing (throwaway) systems		
(Costs for sodium scrubbing systems were not developed in this reference.)	(Costs for sodium scrubbing systems were not developed in this reference.)	● Scrubber is tray tower type.
		● Costs for waste liquor treatment are not included in total capital investment or total annual operating expenses.
		● Sodium alkali is soda ash
Design factors specific to dual alkali systems		
● Scrubbers are perforated plate type.	● Scrubbers are tray type.	● Scrubbers are tray type.
● Solids disposal method is reslurry and ponding of waste solids. Capital costs include thickener, filter, reslurry equipment, 1-mile pipeline for slurry transport, and pond construction.	● On-site ponding.	● Solids disposal method is off-site landfill. Disposal cost is $15.00/ton at 50 percent solids. Capital cost include the costs of a clarifier and filter but not the cost for landfill site preparation.
● Concentrated mode operation with sodium absorbent and lime regenerant.	● Not specified-assumed to be the same as TVA.	● Concentrated mode operation with sodium absorbent and lime regenerant.
Design factors specific to Mag-Ox systems		
● Scrubber is spray grid column type.	● Scrubber is turbulent contact absorber (TCA) type.	(Costs for Mag-Ox systems were not developed in this reference.)
● By-product credit for H_2SO_4 production. Capital and operating costs included for acid plant.	● By-product credit for H_2SO_4 production. Capital and operating costs included for acid plant.	
● Venturi prescrubber for chloride removal included in costs.	● Information on presence of a prescrubber not specified in reference.	
Design factors specific to Wellman-Lord systems		
(Costs for Wellman-Lord systems were not developed in this reference.)	● Scrubber is tray type.	● Scrubber is tray type.
	● By-product credit for H_2SO_4 production. Capital and operating costs included for acid plant.	● By-product credit for sulfur production. Capital and operating costs included for sulfur production facilities.
	● Information on the presence of a prescrubber not specified in reference.	● Venturi prescrubber for chloride removal included in costs.

Figure 5-6. Flue gas desulfurization systems for utility boilers - Total capital investment (March, 1980 dollars).

Costs are for limestone systems. For other technologies multiply costs by:

Lime	0.9
Dual Alkali	1.0
Mag-Ox	1.1
Wellman-Lord	1.0



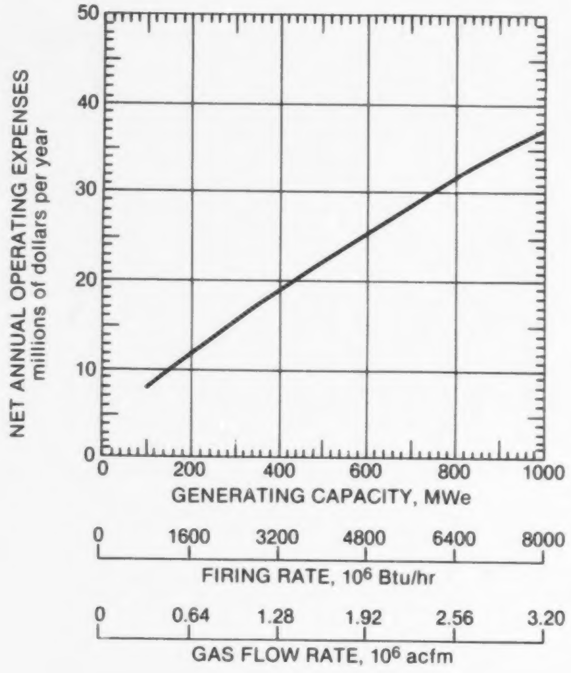
include operating and maintenance costs as well as some capital-investment-related charges. Depreciation is not included. Unit annualized costs are given as cents/kWh and were obtained by calculating annualized cost and dividing by actual annual amount of electricity generated in kwh. Depreciation is included. (See Appendix A for a discussion of annualized cost and unit annualized cost.)

Total capital investment, net annual operating expenses, and unit annualized cost are presented as a function of FGD system capacity. On the graphs for large boilers, the capital cost scale includes the electrical generating capacity (for utility boiler), the boiler firing rate (in 10^6 Btu/hr), and gas flow rate (in 10^3 acfm). Cost for FGD systems on industrial boilers are presented as a function of boiler firing rate (10^6 Btu/hr) and gas flow rate (10^3 acfm).

Figure 5-7. Flue gas desulfurization systems for utility boilers - Net annual operating expenses (March, 1980 dollars).

Costs are for limestone systems. For other technologies multiply costs by:

Lime	0.9
Dual Alkali	1.0
Mag-Ox	1.2
Wellman-Lord	1.0



5.1.3 Major Variables Affecting Costs

The discussion below of some of these variables calls attention to factors which must be considered in evaluating and comparing estimated and reported costs for FGD systems. Utility boiler and industrial boiler FGD system applications are discussed separately in Sections 5.1.4 and 5.1.5, respectively.

5.1.4 Utility Boiler FGD Systems

A review of costs published in the two key sources used in this report for utility FGD systems applied to large boilers showed considerable differences for both total capital investment and net annual operating expenses (2, 5, 6). An analysis of the data reveals that differences are due primarily to design bases rather than errors or inconsistencies in the estimates. Specific differences which illustrate how design criteria have a major impact on costs are described below:

Figure 5-8. Flue gas desulfurization systems for utility boilers - Unit annualized cost (March, 1980 dollars).

Costs are for limestone systems. For other technologies multiply costs by:

Lime	0.9
Dual Alkali	1.0
Mag-Ox	1.2
Wellman-Lord	1.0

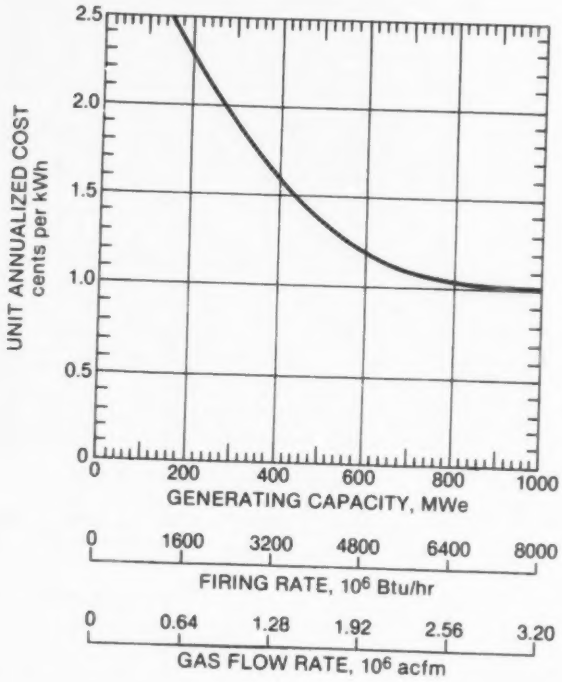
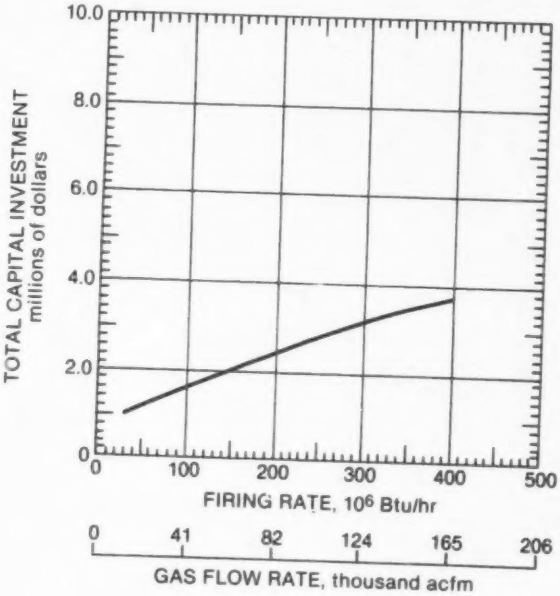


Figure 5-9. Flue gas desulfurization systems for industrial boilers - Total capital investment (March, 1980 dollars).

Costs are for dual alkali systems. For other technologies multiply costs by:

Limestone	1.0
Sodium Alkali	0.7
Wellman-Lord	2.2



- The TVA estimates reflect 78.5 percent sulfur removal, while the PEDCo estimates are based on 90 percent sulfur removal, for coal containing an average of 3.5 percent sulfur. In addition, the PEDCo scrubber design basis accounts for variability from the average in the actual coal sulfur content. The PEDCo scrubber system is designed to achieve 90 percent SO_2 removal from coal which may have an average 3-hour sulfur content as high as 4.61 percent, although the nominal average is 3.5 percent. The TVA estimates do not take into account coal sulfur variability.*
- The TVA design basis does not include redundancy in any process equipment except pumps. Cost increases are substantial when, for example, redundant scrubber modules are employed. The PEDCo cost estimates include a spare scrubbing module and associated equipment, spare pumps, and excess in-process storage capacity.
- Provisions for gas bypass are included in the PEDCo estimates, but not in the TVA estimates.

The effects of the above factors on total capital investment are illustrated in Tables 5-2 and 5-3. Table 5-2 shows the effects of different design bases on the PEDCo-derived total capital investment estimates for several FGD systems. Although the percent reductions in total capital investment given in Table 5-2 are not directly additive, the results are an indication of the size of changes in the total capital investment that can occur by specifying different design bases.

---

*As stated in Section 5.1.2, the published TVA data reflected early design concepts. At the time this report was prepared, new TVA estimates were available only for lime/limestone systems. Costs for other technologies on a consistent TVA basis were not available. The newer TVA estimates for lime/limestone systems are relatively close to the PEDCo estimates. Reported costs for actual systems installed to date are lower that the estimated costs in this report, and lie between the TVA estimates based on early design concepts and the PEDCo based estimates.

Figure 5-10. Flue gas desulfurization systems for industrial boilers - Net annual operating expenses (March, 1980 dollars).

Costs are for dual alkali systems. For other technologies multiply costs by:

Limestone	1.0
Sodium Alkali	0.9
Wellman-Lord	1.2

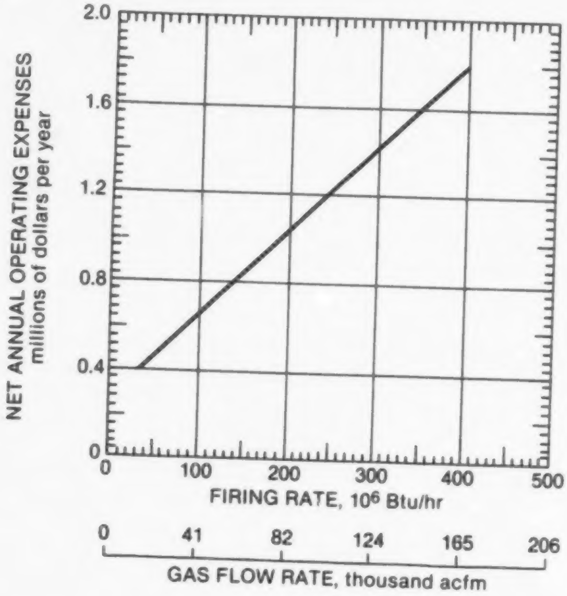


Figure 5-11. Flue gas desulfurization systems for industrial boilers - Unit annualized cost (March, 1980 dollars).

Costs are for dual alkali systems. For other technologies multiply costs by:

Limestone	1.0
Sodium Alkali	0.9
Wellman-Lord	1.5

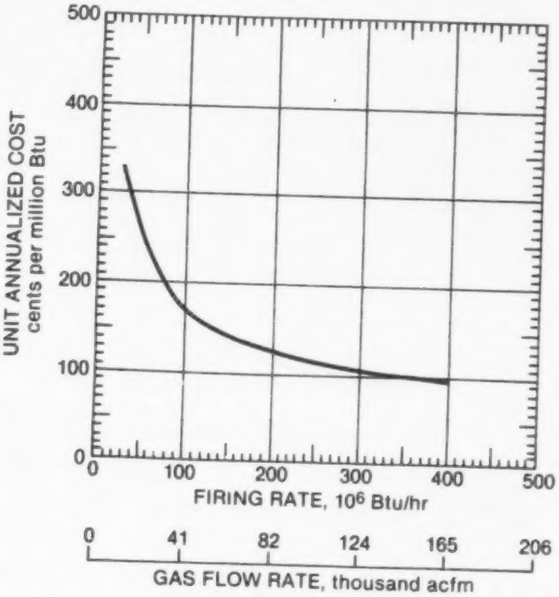


Table 5-3 shows the effects of a modified design basis on the TVA total capital investment estimate for the limestone slurry process. The important result here is the nearly doubled total capital investment figure which results from the modified design basis.

The factors described above, together with other less significant factors in the design bases, account for the difference between the TVA and PEDCo derived total capital investment estimates. At the time this report was initially prepared, TVA was in the process of updating its design basis to include redundancy. Numerous other design basis changes were also being incorporated that would make the TVA and PEDCo design bases more similar. The resulting total capital investment available for a lime/limestone system for the new TVA basis is nearly the same as that given by PEDCo (7).

Actual reported total capital investment for new electric utility FGD systems constructed within the last 10 years falls roughly midway between the TVA (old basis) and PEDCo cost estimates (4). However, when the reported costs are adjusted to include costs for reheat and scrubber redundancy, if not already included in the process configuration, the reported costs are in closer agreement with the PEDCo and the TVA (new basis) estimates. As a result of the 1979 New Source Performance Standards, future utility installations will probably be constructed with redundancy for reliability. For this reason the PEDCo costs were used for the cost curves in this report. A complete set of new TVA costs was not available when this report was being prepared.

Large differences also exist for net annual operating expenses derived from the TVA (2, 5) and PEDCo (6) estimates. Again PEDCo figures result in higher costs. This fact is somewhat misleading, because about 75 percent of these differences are directly related to the higher total capital investment figure used in the PEDCo estimates. The remaining differences are the result of a variety of factors, including sludge handling and fixation chemical costs and higher utility consumption rates.

Differences between the TVA and PEDCo design bases which affect costs have been described above. There are, however, several other factors which can have a significant effect on costs, including:

Table 5-2. Effect of Changes in the Design Basis on PEDCo Total Capital Investment Estimates (6)[a]

Example	Lime process	Limestone process	Dual alkali process	Mag-Ox process	Wellman-Lord process
			Base case capital investment		
Base case - New 500 MWe unit burning 3.5 percent sulfur coal. 90 percent removal based on 3-hour average coal sulfur variability. Five scrubber trains (including one redundant module).	$83,770,000 ($167.5/kWe)	$96,200,000 ($192.4/kWe)	$98,010,000 ($196.0/kWe)	$104,760,000 ($209.5/kWe)	$92,960,000 ($185.9/kWe)
			Percent decrease in total capital investment from base case[b]		
Case I - Emissions level of 1.2 lb $SO_2/10^6$ Btu (~80 percent removal)	10.4% ($150.1/kWe)	11.0% ($171.2/kWe)	10.2% ($176.0/kWe)	10.2% ($188.1/kWe)	8.9% ($169.4/kWe)
Case II - 90 percent removal based on 1-year average coal sulfur variability	4.3% ($160.3/kWe)	N/A[c]	N/A	N/A	N/A
Case III - Elimination of redundant scrubber module	18.8% ($136.0/kWe)	N/A	N/A	N/A	17.4% ($153.6/kWe)

[a] March 1980 dollars.

[b] Example for the lime process - If an emissions level of 1.2 lb $SO_2/10^6$ Btu is used as the design basis, the total capital investment would be $150.1/kWe which is 10.4 percent lower than the base case total capital investment of $167.5/kWe.

[c] N/A - Data not available.

Table 5-3. Effect of Changes in the Design Basis on TVA Total Capital Investment Estimates (5)[a]

Base case - Limestone slurry process, new 500 MWe unit burning 3.5 percent sulfur coal. SO_2 level of 1.2 lb. No scrubber redundancy	$/kWe	Percent increase over base case
Base case total capital investment, $/kWe	107.7	
Modified case - limestone slurry process. New 500 MWe unit burning 6 percent sulfur coal. SO_2 efficiency of 90 percent. 50 percent redundancy.		
Total capital investment increase due to:		
Increased raw material handling	20.2	18.8
Larger waste disposal area and pond	51.8	48.1
50 percent redundancy of ball mills, scrubbers, and other equipment	34.0	31.6
Total increase in total capital investment	106.0	98.5
Modified case total capital investment	213.7	

[a] March 1980 dollars.

- Solid waste disposal method.
- New unit versus retrofit applications.
- Boiler fuel type and sulfur content.

Costs for the lime, limestone, and dual alkali processes are signficantly affected by the choice of the solids disposal method. Capital costs for ponding scrubber slurry are higher than those for landfill disposal of solid waste. Available data (8) suggest that use of landfill disposal reduces total capital investment by up to about 18 percent, depending on the landfill method. However, net annual operating expenses are higher for landfill disposal due to solids handling and disposal costs and raw material costs if stabilization or chemical fixation is required. The choice of the most appropriate waste disposal method for utility boilers is largely dependent on site-specific factors such as land availability, topography, groundwater characteristics, and climate. The total numbers of applications of ponding and landfill disposal methods are about equal. This analysis included both operational and planned FGD systems.

Fuel type and sulfur content also affect both capital investment and annual expenses. Increases in fuel sulfur content result in more sludge and hence a larger solids handling system and waste disposal area for non-regenerable systems. Magnesium oxide and Wellman-Lord systems require larger regeneration facilities. Annual expenses also increase because of higher raw material and utilities consumption and higher waste disposal costs. Table 5-4 shows the effect of coal sulfur content on total capital investment and total net operating expenses (6). Because of lower scrubber costs, FGD systems on oil-fired units cost slightly less than those on fired boilers, there is a smaller volume of gas flow.

5.1.5 Industrial Boiler FGD Systems

The process configuration and costs for industrial boiler FGD systems differ from those of utility boiler FGD systems for a variety of reasons. Some of the more important differences include:

- Waste disposal method.
- Stack gas reheat provision.
- Redundancy.
- Shop-fabricated versus field-erected equipment.

Ponding and landfill waste disposal options are used in equal numbers in applications for lime, limestone, and dual alkali processes on utility boiler FGD systems. Industrial boiler FGD systems, however, generally use the landfill disposal method. Many industrial locations do not have sufficient land available for pond construction. The use of landfill disposal also requires a lower total capital investment.

Stack gas reheat and redundancy in process equipment, often incorporated in utility applications, are not extensively used in industrial boiler FGD systems. The omission of these equipment items

Table 5-4. Effect of Coal Sulfur Content on Total Capital Investment and Total Annual Operating Expenses for Utility Boiler Applications (6)[a, b]

	Lime process	Limestone process	Double alkali process	Mag-Ox process	Wellman-Lord process
Base case - New 500 MWe unit burning 3.5 percent sulfur coal. 90 percent SO_2 removal.					
Base case total capital investment, $	87,770,000	96,200,000	98,010,000	104,760,000	92,960,000
Base case net annual operating expenses, $/yr	23,180,000	24,360,000	28,990,000	29,750,000	22,720,000
Case I - 7.0 percent sulfur coal					
Percent increase in total capital investment	12.7	16.5	12.2	17.9	9.2
Percent increase in net annual operating expenses	26.7	27.2	34.2	25.8	8.6
Case II - 0.8 percent sulfur coal					
Percent decrease in total capital investment	14.4	14.3	N/A[c]	N/A	N/A
Percent decrease in net annual operating expenses	31.9	23.7	N/A	N/A	NA

[a] March 1980 dollars

[b] Example for the lime process - If coal with a sulfur content of 7 percent is used as the design basis, the total capital investment would increase by 12.7 percent over the base case total capital investment.

[c] N/A - data not available

substantially reduces the required total capital investment. The costs presented in this report for industrial boiler FGD systems do not include these items.

An obvious difference between utility boiler and industrial boiler FGD systems is their relative size. An important consequence of this size difference is the manner in which the FGD systems are constructed. Because of their size, utility systems must be field-erected, while major components of the smaller industrial boiler FGD systems can be shop-fabricated and transported by truck or rail to the site. Total capital investment is lower for shop-fabricated equipment than for field-erected equipment.

Total capital investment and net annual operating expenses for industrial boiler FGD systems are also affected by the required percent SO_2 removal and the fuel sulfur content. The effect of these variables on capital investment and operating expenses is summarized in Table 5-5.

There are also differences between coal- and oil-fired boilers. Coal-fired boilers typically result in a higher flue gas rate for a given firing rate than oil-fired boilers. This is due to combustion characteristics of the different fuels. As a result, the scrubbing section of an FGD system is more expensive for coal-fired applications at comparable firing rates.

5.2 Lime Spray Drying Process

Spray drying is a relatively new FGD technology. At this writing, three industrial systems and one large-scale (100 MWe) utility demonstration system are operational. There are another four industrial-size spray drying systems planned or under construction (9). No commercial utility lime spray drying systems are operating. However, 10 utilities have purchased such systems, and about half of those are scheduled to start up in the next 2 or 3 years.

In spray drying systems the SO_2 gas is either absorbed or adsorbed onto the sprayed materials.

The utility systems will be used on boilers firing low sulfur coal (1.5 percent sulfur or less) in most cases. The sulfur content of the coal burned in the industrial applications ranges from 0.7 to 3.5 percent in the operating or planned systems to date.

5.2.1 Process Description

In lime spray drying systems, flue gas at air preheater outlet temperatures (generally between 250 and 350°F) is contacted with a finely atomized lime slurry in a spray dryer. Figure 5-12 is a block flow diagram of the spray drying process. The flue gas is adiabatically humidified to within 20 to 50°F of its saturation temperature as water evaporates from the slurry. SO_2 in the flue gas reacts with the calcium hydroxide in the slurry to form calcium sulfite, some of which is oxidized to calcium sulfate by oxygen in the flue gas. Heat from the flue gas dries the calcium sulfite and sulfate solids to less than 1 percent residual moisture. The bulk of these solids, along with the fly ash in the flue gas, pass through the dryer and are collected in a downstream fabric filter or electrostatic precipitator (ESP). In some system designs, a portion of the solids are collected from the bottom of spray dryer.

Table 5-5. **Effect of Coal Sulfur Content and SO_2 Removal Efficiency on Total Capital Investment and Total Annual Operating Expenses for Industrial Boiler Applications (3)**[a, b]

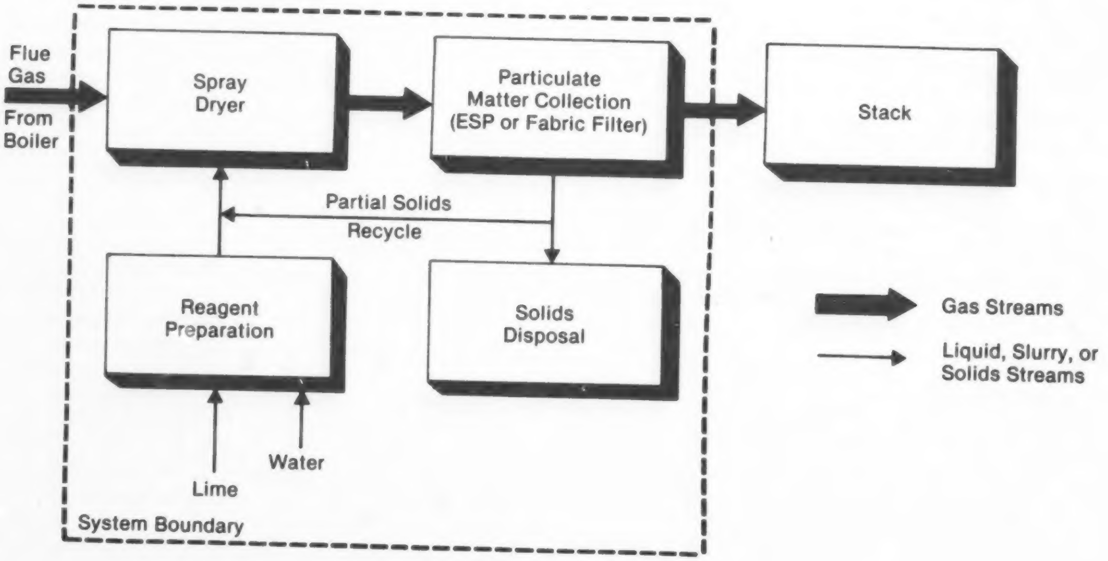
	Limestone process	Sodium throwaway process	Double alkali process	Wellman-Lord process
Base case - New FGD unit applied to a 200×10^6 Btu/hr boiler burning 3.5 percent sulfur coal. 90 percent SO_2 removal.				
Base case total capital investment, $	2,100,000	1,480,000	2,230,000	4,960,000
Base case net annual operating expenses, $/hr	1,060,000	1,000,000	1,050,000	1,420,000
Case I - 75 percent removal				
Percent decrease in total capital investment	7.4	2.3	N/A[c]	N/A
Percent decrease in net annual operating expenses	10.8	7.7	N/A	N/A
Case II - 0.6 percent sulfur coal				
Percent decrease in total capital investment	15.3	16.1	16.1	45.4
Percent decrease in net annual operating expenses	33.0	41.2	34.9	41.0

[a] March 1980 dollars

[b] Example for limestone process - If an SO_2 removal efficiency of 75 percent is used as the design basis, the total capital investment would be 7.4 percent lower than the base case total capital investment.

[c] N/A - Data not available

Figure 5-12. Lime spray drying process for flue gas desulfurization.


Flue Gas
From Boiler
Spray Dryer
Particulate Matter Collection (ESP or Fabric Filter)
Stack
Partial Solids Recycle
Reagent Preparation
Solids Disposal
Lime
Water
System Boundary
Gas Streams
Liquid, Slurry, or Solids Streams

The dry waste product from the lime spray drying process is usually disposed of by landfill. As shown in Figure 5-12, a portion of the product solids/fly ash mixture can be recycled back to the dryer. This scheme reduces fresh lime requirements by taking advantage of any unreacted reagent or availability of fly ash alkalinity in the solids.

The reader should note that, in contrast to the wet FGD system boundaries presented earlier, the spray drying system *includes* a fabric filter for particulate control.* The particulate matter collection system is an inherent component of the dry FGD system, and it has been observed to contribute to adsorption of additional SO_2. The reader should take this fact into account when comparing the costs for spray drying with those presented earlier for wet FGD systems which do not include the costs of particulate matter control.

Another, less significant difference in the scope of the wet FGD systems and lime spray drying systems is that no stack gas reheat is included in the spray drying system. Unlike wet FGD systems, the spray drying process does not result in saturated flue gas and the need for reheat is reduced.

5.2.2 Design Basis and Costs

Costs presented in this section for utility lime spray drying systems were adapted from estimates developed by TVA (10). Costs for industrial applications were adapted from original estimates developed by Radian Corporation (11). Each of these sources contains estimates representative of recent lime spray drying system design. Table 5-6 lists the major design parameters for both the utility (10) and industrial systems (11).

Costs presented in this section have been developed for lime spray drying systems applied to boilers firing a relatively low sulfur (0.7 percent) coal. This is in contrast to the costs presented earlier for wet FGD systems which were based on a 3.5 percent sulfur coal. The primary reason for limiting the cases examined to the low sulfur coal is the lack of documented information on application of spray drying to high sulfur coal-fired boilers. No utility systems have been sold for high sulfur units (greater than 3 perent sulfur coal). The TVA has developed some estimates for high sulfur utility applications. These costs are discussed in Section 5.2.3.

Total capital investment, net annual operating expenses, and unit annualized cost for utility systems are presented in Figures 5-13, 5-14, and 5-15, respectively. The corresponding costs for industrial boiler applications are presented in Figures 5-16, 5-17, and 5-18.

Table 5-6. Design Bases for Utility and Industrial Lime Spray Dryer/Fabric Filter Systems

Utility systems (10)

- Cost estimates are for new units.
- Costs are based on 70% SO_2 removal (0.6 lb/10^6 Btu controlled emissions) and a low sulfur eastern coal with the following properties: 0.7% S, 15% ash, 11,700 Btu/lb.
- Plant heat rate is 9500 Btu/kWh.
- Costs are derived from TVA 500 MWe case. Curves based on calculations for 200, 500, and 1000 MWe cases.
- Spray dryers have rotary atomizers; 84% of gas is treated in spray dryers at 83% removal, for an overall SO_2 removal of 70%. Partial bypass is used to reheat gas entering fabric filter. Stack gas exit temperature is 175°F.
- Costs for a reverse-air fabric filter (air-to-cloth ratio of 2.5:1) are included.
- Reagent stoichiometry is assumed to be 1.0 mole lime per mole of inlet SO_2.
- System redundancy includes one spare dryer, redundant reagent preparation area, and SO_2 emergency flue gas bypass.
- Costs include all ductwork associated with system.
- Waste solids transported 1 mile to landfill. Landfill assumed to have clay liner. All capital and operating costs for landfill included in system cost.
- 30-year plant life; 5500 hr/yr operation (65 percent operating factor).
- Midwest location.

Industrial systems (11)

- Cost estimates are for new units.
- Costs are based on 70% SO_2 removal and a low sulfur eastern coal with the following properties: 0.7% S; 15% ash; 11,700 Btu/lb.
- Cost curves drawn from calculations for 100, 200, and 400 million Btu/hr cases.
- Spray dryer equipped with rotary atomizer; designed to treat 84% of the flue gas at 83% removal for an overall removal of 70%. Partial bypass used to reheat gas entering fabric filter. Stack gas exit temperature is 175°F.
- Costs for a reverse-air fabric filter (air-to-cloth ratio of 2.0:1) are included.
- System redundancy includes spare pumps.
- Costs include all associated ductwork.
- Waste solids trucked to off-site landfill. Landfill capital and operating costs not included in system (only charge per ton of waste disposed).
- 30-yr plant life; 5256 hr/yr operation (60 percent operating factor).
- Midwest location.

5.2.3 Major Variables Affecting Costs

Costs of spray drying FGD systems, like those for wet FGD presented earlier, are sensitive to a variety of site specific factors. These include utility versus industrial applications, boiler fuel properties, and whether an ESP or fabric filter is used as the solids collecting device in the spray drying system.

The same considerations described in Section 5.1.5 for wet FGD systems apply to the differences between industrial and utility spray drying systems:

- industrial systems are likely to use off-site landfill for waste disposal rather than on-site ponding or landfilling;

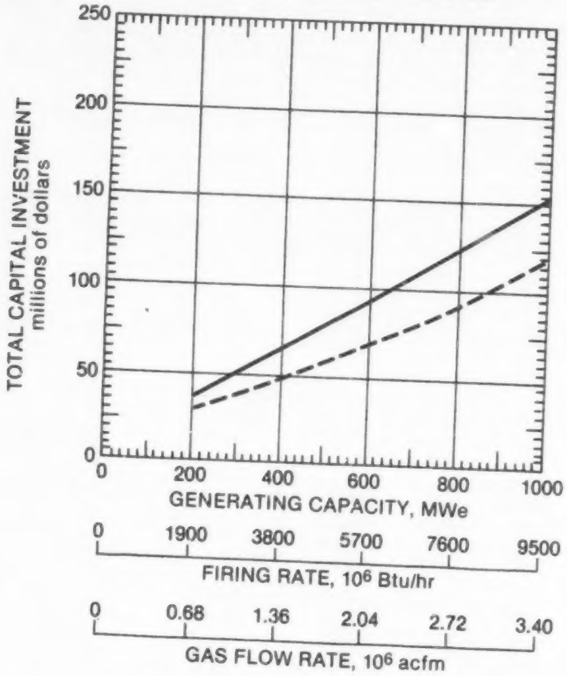
---

*Some commercial spray drying systems could include an ESP instead of a fabric filter, depending on site specific factors. However, a fabric filter has been selected here since only one system sold to date has included an ESP.

Figure 5-13. Lime spray drying flue gas desulfurization systems for utility boilers - Total capital investment (March, 1980 dollars).

Coal sulfur content, 3.5%; 90% removal ———
Gas flow rate for 3.5% S coal is 0.82 times the rate shown on the scale below.

Coal sulfur content, 0.7%; 70% removal - - - - -



- industrial systems generally do not have the extensive redundancy or stack gas reheat provisions that are typically included with utility systems; and
- industrial system equipment is likely to be shop-fabricated rather than field-erected.

For spray drying/fabric filter systems another difference may be in the type of fabric filter selected. Due primarily to pressure drop and bag-wear considerations, utility systems will generally have a reverse-air fabric filter. However, in industrial systems where pressure drop considerations are not so critical, a pulse-jet fabric filter is sometimes used. The capital costs of a pulse-jet fabric filter are less than a reverse-air unit of the same capacity.* (See Section 4.3 in this report.)

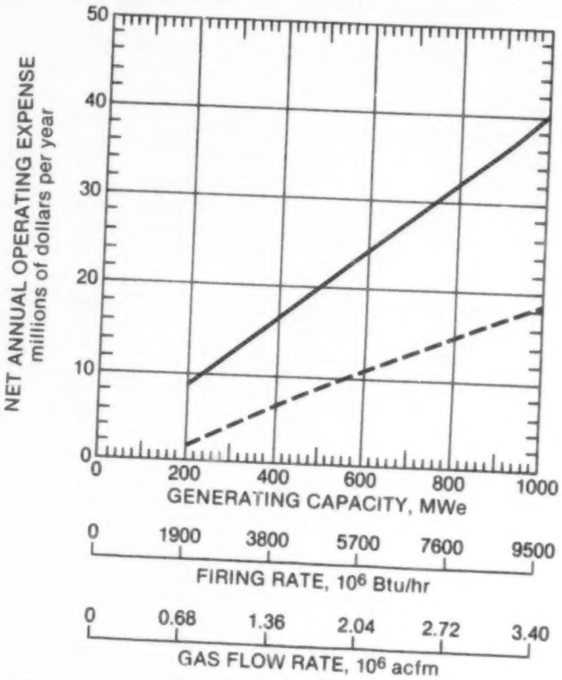
The fuel properties that affect spray drying system costs are:

- Heating value and moisture content.
- Fuel sulfur content.
- Fuel ash content.
- Available alkalinity in the fly ash.

Figure 5-14. Lime spray drying flue gas desulfurization systems for utility boilers - Net annual operating expenses (March, 1980 dollars).

Coal sulfur content, 3.5%; 90% removal ———
Gas flow rate for 3.5% S coal is 0.82 times the rate shown on the scale below.

Coal sulfur content, 0.7%; 70% removal - - - - -



Heating value and moisture content, as well as the excess air rate to the boiler, impact the amount of flue gas that must be treated in the spray drying system. Combustion of low rank western coals or lignite can result in about 10 to 30 percent higher flue gas flows (10) with a corresponding increase in the size and cost of the spray drying and solids removal equipment.

The fuel sulfur content also affects costs of the spray drying systems. A higher fuel sulfur content results in a direct increase in fresh lime requirements and also increases the amount of waste solids that must be disposed of. In addition, the stoichiometric ratio of fresh reagent to inlet SO_2 to achieve a given removal increases as the inlet SO_2 concentration increases (12). Thus an increased fuel sulfur content will require a larger feed handling and preparation system and landfill area and will result in increased annual reagent and waste disposal costs.

Figures 5-13 through 5-15 show a comparison of costs for lime spray drying systems applied to high and low sulfur coal utility boilers.† The high sulfur

---

*Reverse-air fabric filters are included in the industrial spray drying system costs presented in this section, representing a conservative design practice.

†The high sulfur coal system includes solids recycle, but no gas bypass for reheat.

Figure 5-15. Lime spray drying flue gas desulfurization systems for utility boilers - Unit annualized cost (March, 1980 dollars).

Coal sulfur content, 3.5%; 90% removal ———
Gas flow rate for 3.5% S coal is 0.82 times the rate shown on the scale below.

Coal sulfur content, 0.7%; 70% removal - - - - -

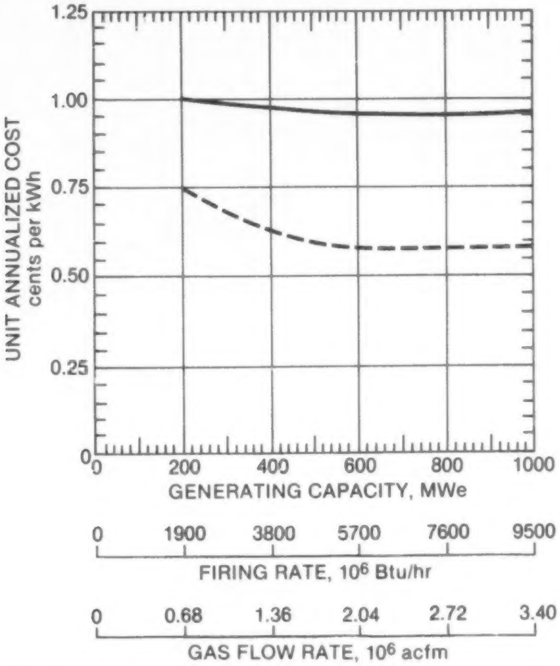
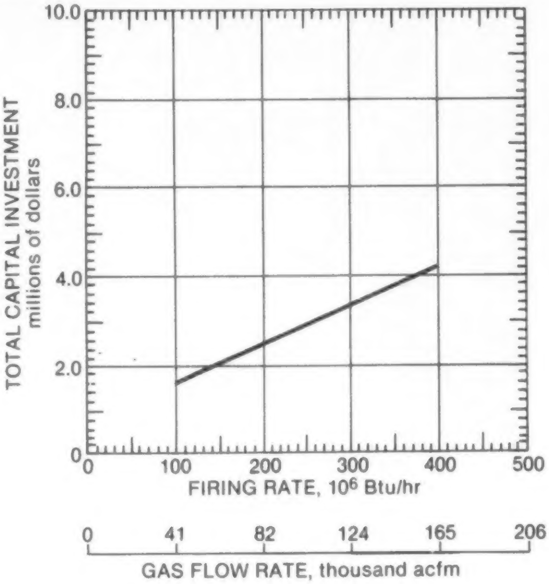


Figure 5-16. Lime spray drying flue gas desulfurization systems for industrial boilers - Total capital investment (March, 1980 dollars).

Coal sulfur content, 0.7%; 70% removal



case is based on a lime-to-SO_2 stoichiometry of 1.6 and 90 percent removal; whereas, the low sulfur case costs are based on a 1.1 stoichiometry and 70 percent SO_2 removal.

Stoichiometric requirements for high SO_2 removal in high sulfur coal applications are not well-documented. No utility systems have been sold for high sulfur coal units. And, although at least two industrial spray drying systems treating high sulfur flue gas have been sold, no data on the stoichiometric requirements have been reported. Thus, there is substantial uncertainty regarding reagent-related costs for high sulfur applications of spray drying.

The TVA estimates that a 20 percent increase in lime stoichiometry would result in about a 7.5 percent increase in unit annualized cost for a high sulfur coal application, and about a 2 percent increase for low sulfur applications (10).

Waste solids collected from the system include fly ash along with waste solids from the spray dryer. The volume of waste to be disposed of and capital and operating costs associated with landfill are thus increased as the ash content (fly ash emissions) from the coal is increased.

The availability of alkaline species in the fly ash to react with SO_2 in the spray dryer can substantially reduce fresh lime requirements (13). One method used to take advantage of the fly ash alkalinity is to recycle some of the waste solid/fly ash mixture back to the spray dryer. This operating method also results in recycle of unreacted lime in the waste solids. Although solids recycle is not included in the costs presented here, there are cases where the cost of the recycle equipment can be offset by the resulting reduced costs for lime. Specifically, applications with high lime stoichiometric requirements (high sulfur applications) or those in which the fly ash has high available alkalinity are instances in which solids recycle may be of significant benefit.

An ESP can be used in a spray drying system instead of a fabric filter. The choice depends on user preference and site-specific factors such as fly ash resistivity, ESP inlet dust loading, and pressure drop considerations. The ESP may or may not be more economical in some cases than the fabric filter; the comparison is highly site-specific.

Figure 5-17. Lime spray drying flue gas desulfurization systems for industrial boilers - Net annual operating expenses (March, 1980 dollars).

Coal sulfur content, 0.7%; 70% removal

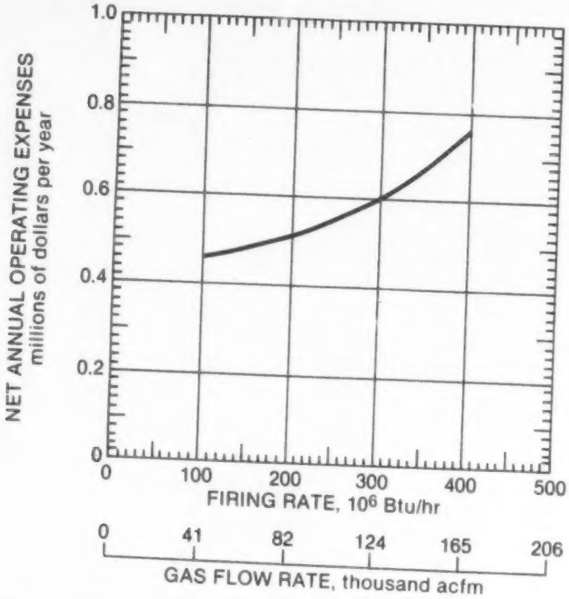
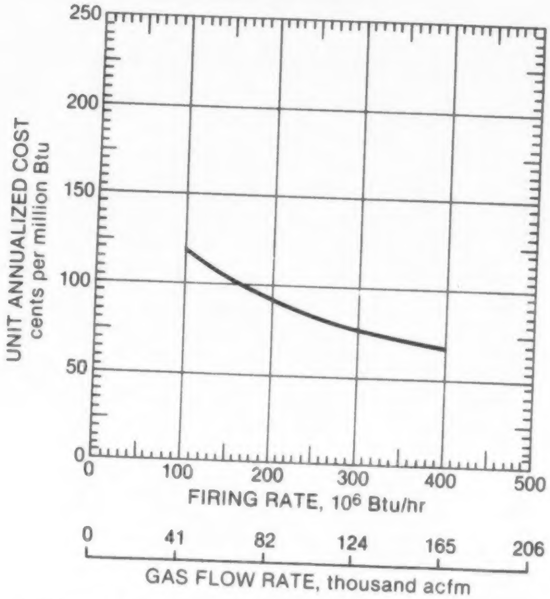


Figure 5-18. Lime spray drying flue gas desulfurization systems for industrial boilers - Unit annualized cost (March, 1980 dollars).

Coal sulfur content, 0.7%; 70% removal



References - Section 5

Wet Scrubbing Systems

1. U.S. Environmental Protection Agency. Appendix A: Compilation of Air Pollutant Emission Factors, AP-42. Office of Air Quality Planning and Standards, Research Triangle Park, NC. February 1980. p. A-2.
2. Anderson, K.D., et al. (TVA) Definitive SO_x Control Process Evaluations: Limestone, Lime, and Magnesia FGD Processes, EPA-600/7-80-001, PB80-196314*. Prepared for U.S. Environmental Protection Agency, Washington, DC, January 1980, 296 pp.
3. Dickerman, J.C., and K.L. Johnson. (Radian Corp.) Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization, EPA-600/7-79-178i, PB80-150873*. Prepared for U.S. Environmental Protection Agency, Washington, DC, November 1979.
4. Smith, M., et al. (PEDCo Environmental, Inc.) EPA Utility FGD Survey: April-June 1980, EPA-600/7-80-029c, PB80-226335*. Prepared for U.S. Environmental Protection Agency, Washington, DC, July 1980, Appendix A, pp. A-1 through A-25.
5. Tomlinson, S.V., et al. (TVA) Definitive SO_x Control Process Evaluations: Limestone, Double Alkali, and Citrate FGD Processes, EPA-600/7-79-177, PB80-105828*. Prepared for U.S. Environmental Protection Agency, Washington, DC, August 1979, 236 pp.
6. Gibbs, L.L. (PEDCo Environmental, Inc.) Particulate and Sulfur Dioxide Emission Control Costs for Large Coal-Fired Boilers, EPA-450/3-78-007, PB 281 271. Prepared for U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards, Research Triangle Park, NC, February 1978, 168 pp.
7. McGlamery, G.G., et al. (TVA) FGD Economics in 1980. In Proceedings: Symposium on Flue Gas Desulfurization - Houston, October 1980; Vol. 1, EPA-600/9-81-019a, PB 81-243 156. April 1981, pp. 49-83.
8. Barrier, J.W. et al. (TVA) Economics of Disposal of Lime/Limestone Scrubbing Wastes: Sludge/Fly Ash Blending and Gypsum Systems, EPA-600/7-79-069, PB 297 946*. Prepared for U.S. Environmental Protection Agency, Washington, DC, February 1979, p. xxvii.

Spray Drying Systems

9. Kelly, M.E., and S.A. Shareff. (Radian Corp.) Third Survey of Dry SO_2 Control Systems. EPA-600/7-81-097, PB 81-218976*, Prepared for U.S. Environmental Protection Agency, Research Triangle Park, NC, June 1981. p. 15-16.
10. Burnett, T.A., and K.D. Anderson. (TVA) Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD systems. EPA-

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

600/7-81-014, PB81-206476*. Prepared for U.S. Environmental Protection Agency, Research Triangle Park, NC, February 1981. p. 91-235.

11. Jennings, M.S., and M.E. Kelly. (Radian Corp.) Costs of Sulfur Oxide and Particulate Matter Emission Control for Coal- and Oil-Fired Industrial Boilers. EPA Contract 68-02-3058; ESED Project 76/13. Prepared for U.S. Environmental Pretection Agency, Research Triangle Park, NC, August 21, 1981. p. 2-1 to 2-32.
12. Apple, C. and M.E. Kelly. (Radian Corp.) Mechanisms of Dry SO_2 Control Processes. EPA-600/7-82-026, PB 82-196924*. Prepared for U.S. Environmental Protection Agency, Research Triangle Park, NC, April 1982. p. 44-53.
13. Gibson, E.D., M.A. Palazzolo, and M.E. Kelly. (Radian Corp.) Summary Report: Sulfur Oxides Control Technology Series: Flue Gas Desulfurization Spray Drying Process. EPA Contract 68-02-3171, Task 37. Prepared for U.S. Environmental Protection Agency, Research Triangle Park NC, September 15, 1981.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

Appendix A
Methods for Adjusting Data

The cost data in this report were derived from cost data in existing published sources. These data were adjusted to conform to the format used in this report, to reflect total system costs rather than individual system component costs, and to update costs from a variety of price years to the common reference time of March 1980. The exact method used for adjusting data varied depending on the form of the original data. For some technologies, well documented costs for total systems were already available. For other technologies, estimates had to be prepared from individual component costs that were available in the literature. This varied between the four major technology areas of this report as well as between individual technologies within a technology area. The following sections describe the format, cost factors, and unit prices used to develop system costs; methods and cost indices to update costs; and special considerations in using published cost data by technology area.

A.1 Format, Cost Factors, and Unit Prices

The format for presenting all cost data in this report is based on an earlier report by Uhl(1). All capital investment costs are presented as total capital investment, and annual costs are presented both as net annual operating expenses and unit annualized cost. Table A-1 defines the cost elements comprising total capital investment as used here. Table A-2 defines net annual operating expenses. For a detailed discussion of cost elements the reader is referred to the report by Uhl(1). Unit annualized cost is derived from net annual operating expenses as explained in Section A.2

In addition to listing the cost elements, the tables also contain cost element item numbers assigned to those line items in the Uhl report, as well as cost factors used in the present work to derive the various cost elements from preceding line items by factoring as discussed next.

The computation of total capital investment as shown in Table A-1 begins with the total direct cost for the system under consideration. This total direct cost is the total direct installed cost of all capital equipment comprising the system. In some references, especially for water and wastewater treatment systems, this cost is referred to as total construction cost. Some authors use other names for this line item. Depending on the reference and the technology, the direct capital cost was available or was derived from uninstalled equipment costs by computing costs of installation separately. Literature costs were updated to March 1980 using capital cost indices. These indices are discussed in Section A.3. To obtain the total capital investment, other costs must be added to the total direct cost. A standard procedure of cost estimating is to obtain these other costs by factoring.

The first group of other cost elements is indirect costs. These include engineering and supervision, field construction expenses, and various other expenses such as general project administration and legal fees, for example. These costs are computed by multiplying total direct costs by a factor as shown in Table A-1. The factor is approximate, is obtained from the cost literature, and is based on previous experience with capital projects of a similar nature. This is true also for the factors for other cost elements shown in the table. Factors can have a range of values and vary according to technology area and for individual technologies within an area. Appropriate factors were selected for use in this report based on the authors' judgment and experience.

When the indirect costs are added to the total direct costs, total bare module cost is obtained. The cost elements in the next group are obtained by applying factors to the total bare module cost. These cost elements are added to the total bare module cost to obtain total plant cost. Some additional cost elements can be calculated from the total plant cost by applying factors. These additional cost elements can include interest during construction and start-up costs if these costs are included as part of the total capital investment.* If these costs are capitalized, as they are in this report, they are added to the total plant cost to obtain the total depreciable investment (sometimes referred to as total fixed capital as well as other names). The total depreciable investment is used in calculating the unit annualized cost discussed in Section A.2.

---

*This is an option that depends on accounting practices of individual organizations. If these costs are not capitalized they are treated as an expense, in the first year of operation, for example.

Table A-1. Format and Factors for Total Capital Investment[a]

Item No.[b]	Item	Drinking water	Wastewater treatment	Particulate matter control[c] MC	ESP	FF	WS	Flue gas desulfurization Wet	Dry
1-10	Direct cost items[d]	individual cost items vary widely from technology to technology							
11	TOTAL DIRECT COST	Sum of 1-10	Sum of 1-10	Sum of 1-10				Sum of 1-10	Sum of 1-10
12-20	Indirect cost items (Engineering and supervision, construction and field expenses, other)	0.15	0.15	0.30	0.20	0.15	0.30	0.24	0.24
21	TOTAL BARE MODULE COST	Sum of 11-20	Sum of 11-20	Sum of 11-20				Sum of 11-20	Sum of 11-20
22	Contingency	0.15	0.15	0.10	0.10	0.10	0.10	0.21	0.21
23	Contractor's fee	0.10	0.10	0.10	0.05	0.05	0.10	0.04	0.04
27	Retrofit increment	-	-	-				-	-
24-26, 28-30	Other	-	-	-				-	-
31	TOTAL PLANT COST	Sum of 21-30	Sum of 21-30	Sum of 21-30				Sum of 21-30	Sum of 21-30
32	Interest during construction[e]	0.12	0.12	0.12	0.03	0.03	0.12	0.19	0.19
33	Start-up	0.05	0.05	0.05	0.01	0.01	0.05	0.08	0.08
34	Other	-	-	-				-	-
35	TOTAL DEPRECIABLE INVESTMENT	Sum of 31-34	Sum of 31-34	Sum of 31-34				Sum of 31-35	Sum of 31-35
36	Land	$2000/acre	$2000/acre	$2000/acre				$2000/acre	
37	Working capital	0.10	0.10	0.10	-[f]	-[f]	0.10	-[f]	-[f]
38-40	Other	-	-	-				-	-
41	TOTAL CAPITAL INVESTMENT	Sum of 35-40	Sum of 35-40	Sum of 35-40				Sum of 35-40	Sum of 35-40

[a] The capitalized line items are the subtotals of the preceding line items including the preceding subtotal. For example, TOTAL BARE MODULE COST is the sum of TOTAL DIRECT COST and indirect cost items. Factors shown are multiplied by the preceding subtotal line item to obtain the line items to which the factors correspond. For example, the factor for contingency under Drinking Water is 0.15. This factor is multiplied by TOTAL BARE MODULE COST to obtain the contingency line item for drinking water systems.

[b] Refers to line item code proposed in report by Uhl (1).

[c] MC = mechanical collectors; ESP = electrostatic precipitators; FF = fabric filters; WS = wet scrubbers.

[d] Direct cost items are installed equipment costs including such components as piping, insulation, electrical work, instrumentation, concrete, and structural steel.

[e] This interest factor reflects the longer construction period for FGD systems than for the other environmental technology areas.

[f] Working capital for ESP, FF, and FGD systems was computed as 25% of processing expenses rather than as a percentage of TOTAL DEPRECIABLE INVESTMENT.

Finally, the capital requirements for land and working capital are added to the total depreciable investment to obtain total capital investment. In this report estimated land requirements and a unit price of $2000 per acre were used to calculate the land cost for each technology. Working capital can be computed in a number of different ways. Here it was estimated as a percentage of total depreciable investment except for ESP, FF, and FGD systems where it was calculated as a percentage of processing expenses.

Cost elements for net annual operating expenses are shown in Table A-2. Direct cost elements are added together to yield processing expenses. Some references refer to these expenses as operating and maintenance costs. Values for these cost elements were obtained from the literature and updated using unit prices for March 1980 given in Table A-3. This was usually accomplished by ratioing new unit prices to the old and multiplying by the reported annual value for that cost element. In some cases the annual value for a cost element was calculated directly from the operating requirement (e.g., labor hours per year) multiplied by the unit price. The method used depended on how data were presented in the literature source. Overhead was calculated as a fraction of labor costs. Insurance, property taxes, and general expenses were calculated as a fraction of total depreciable investment.

A.2 Unit Annualized Cost Calculations

Unit annualized cost is derived from net annual operating expenses and capital changes as discussed in the next paragraph. The annualized cost is first calculated and then divided by system capacity to

Table A-2. Format for Net Annual Operating Expenses

Item No.	Item	Explanation
53	Raw materials	Computed as annual operating requirement multiplied by unit price.
56-58, 61	Labor[a]	
59, 60	Materials[a]	
62	Steam	
63	Power (Electricity)	
64	Compressed air	
65	Water	
66	Fuel	
67	Waste disposal	
68, 69	Other	
70	PROCESSING EXPENSES	Sum of items 53-69 (except 54, 55)
74	Overhead[b]	50% of labor; 65% of labor for ESP, FF[c]
76	Insurance and property taxes	1% of TOTAL DEPRECIABLE INVESTMENT (35)
80	NET OPERATING COSTS	Sum of items 70, 74, & 76
87	General expense	4% of TOTAL DEPRECIABLE INVESTMENT (35)
88-89	Other	
90	NET ANNUAL OPERATING EXPENSES	Sum of items 80-89 (except 81-86)

[a] Costs for maintenance labor and materials are expressed separately in the labor and materials categories.

[b] For FGD systems: 60% of processing expenses less utilities.

[c] ESP = electrostatic precipitator; FF = fabric filter.

Table A-3. Unit Prices Employed for Net Annual Operating Expenses[a]

1. Direct labor rate, \$/hr	= 11.40
2. Energy costs	
Electric power, \$/kWh	= 0.04
Fuel oil, \$/gal.	= 0.60
Natural gas, \$/10^6 Btu	= 2.12
Gasoline, \$/gal.	= 1.23
3. Land, \$/acre	= 2000
4. Chemical costs	
Chlorine, \$/ton	= 300
Limestone, \$/ton	= 8
Agricultural limestone, \$/ton	= 15
Lime, \$/ton	= 40
Soda Ash, \$/ton	= 90
Magnesium oxide, \$/ton	= 300
Ferric chloride, \$/ton	= 100
Alum, \$/ton	= 72
Sodium chlorite, \$/ton	= 97
Ammonia:	
Anhydrous, \$/ton	= 130
Aqueous, \$/ton	= 175
Activated carbon (granulated), \$/lb	= 0.50
Sulfuric acid (credit for FGD systems), \$/ton	= 25.00
Catalyst (Mag-Ox FGD process), \$/liter	= 2.50
Fixation chemicals (FGD L/LS/DA processes), \$/ton	= 20
5. Other unit costs	
Process water, \$/1000 gal.	= 0.12
Steam, \$/10^6 Btu	= 2.00
Waste disposal (sludge handling utility FGD systems), \$/ton-mile	= 2.00
Waste disposal (sludge handling industrial boiler FGD systems and particulate matter technologies), \$/ton	= 15.00

[a] Applicable to March 1980.

yield cost per unit of capacity. Conversion factors are applied as necessary to express the result in appropriate units. For example, the annualized cost for a wastewater technology expressed as millions of dollars per year is divided by system capacity in millions of gallons per day and adjusted with appropriate conversion factors to obtain the unit annualized cost in cents per thousand gallons.

The annualized cost corresponds to a uniform annual revenue requirement to cover both net annual operating expenses as well as capital recovery, return on investment, and income taxes. The capital recovery, return on investment, and resulting income tax requirements are directly related to the capital investment and are referred to as capital charges.* The net annual operating expense cost elements of insurance, property taxes, and general expenses, when derived from capital investment by factoring as they were in this report, can also be considered capital charges. Because the capital investment is fixed, these capital charges are also referred to as fixed charges. Income tax is included with capital charges as a fixed charge. The remaining cost elements of net annual operating expenses, such as labor, are variable because they can change with the level of operation of the system, and can be referred to as variable charges. The annualized cost is, therefore, the sum of the variable and fixed annual charges for the technology.

The fixed annual charges can be computed by different methods. Here the fixed charge rate method is related to methods employed by the Electric Power Research Institute (EPRI), the Jet Propulsion Laboratory (JPL), and the Mitre Division of the Mitre Corporation (2, 3, 4). Doane et al. (3) discusses the procedure by JPL for calculating fixed charge rate FCR, which has been recommended for use in this report by Uhl (5).

The fixed charge rate is multiplied by the total depreciable investment to obtain the fixed annual charges as a single number. The fixed annual charges are added to the variable annual charges, as discussed above, to obtain annualized cost. In equation form,

$$AC = FCR \times TDI + VAC$$

where AC = annualized cost
FCR = fixed charge rate
TDI = total depreciable investment, and
VAC = variable annual charges.

The unit annualized cost is then,

$$UA = AC/CAP$$

where UA = unit annualized cost, and
CAP = system operating capacity.

Based on the discussion by Doane et al. (3), the FCR is found from,

$$FCR = CRF_{k,N}\left[\frac{1-\tau \times DPF_{m,k,n}-\alpha}{1-\tau}\right]+\beta_1+\beta_2+\beta_3$$

---

*Capital recovery and return on investment are otfen expressed in a numerically equivalent form as depreciation and interest. Preferred terminology depends on perspective; whether the charges are viewed from an investor's, borrower's, or lender's viewpoint.

where $CRF_{k,N}$ = the capital recovery factor computed at cost of capital k over N years

$CRF_{k,n}$ = the capital recovery factor computed at cost of capital k over n years

k = after-tax cost of capital or internal rate of return

n = taxable life

N = system book lifetime

τ = income tax rate expressed as a decimal fraction

$DPF_{m,k,n}$ = depreciation factor for m-type depreciation, at an after-tax cost of capital, k, over n years accounting or taxable lifetime

α = investment tax credit, and

$\beta_i (i = 1,2,3)$ = miscellaneous fixed charges of insurance, property taxes, general and administrative expenses.

The after-tax cost of capital is found from,

$$k = (1 - \tau)\, k_d \frac{D}{V} + k_c \frac{C}{V} + k_p \frac{P}{V}$$

where k_d = the cost of debt capital

k_c = common stockholder's rate of return on investment and therefore the cost of common equity capital

k_p = preferred stockholder's rate of return on investment and therefore the cost of preferred equity capital

D/V = ratio of debt to total capitalization

C/V = ratio of common stock to total capitalization, and

P/V = ratio of preferred stock to total capitalization.

One of the several equivalent algebraic forms for capital recovery factor is,

$$CRF_{k,N} = \frac{k}{1 - (1 + k)^{-N}}$$

The depreciation factor $DPF_{m,k,l}$ depends on m, the depreciation method used (e.g., straight line, sum-of-the-years digit, or others). When straight line depreciation is used, as was done in this report, the factor is,

$$DPF_{sl,k,n} = \left[n \cdot CRF_{k,n} \right]^{-1}$$

The financial premises used in calculating the fixed charge rate for each technology in this report are given in Table A-4. Also given are the resulting values for weighted after-tax cost of capital and several terms (including the first) for the fixed charge rate equations. Calculations are based on a system book lifetime of 20 years and a tax life of 10 years.

The basis for these assumptions is as follows. Interest on debt financing of 13 percent was assumed typical of rates paid by private companies in non-regulated industries* and 14 percent by regulated industries (e.g., utilities). For municipal projects, 10 percent reflects investor acceptance of a lower rate of return on tax-free bonds than for taxable corporate debt instruments. For non-regulated private industry, a common stockholder return of 14 percent was assumed, which reflects a combined typical common stock dividend rate and an expected capital gain. For regulated private industry, such as electric utilities, a total equity return of 12 percent was assumed to reflect the higher dividend rate and lower overall risk associated with utilities.

The financing mix assumed for each technology reflects the most likely use of that technology by economic sector. Drinking water systems are usually public projects financed by bonded debt. Wastewater projects can be either public or private. For public projects, all financing was assumed to be by bonded debt. For private companies the assumption was 25 percent financing by debt and 75 percent by equity. Particulate control would most likely be used by either non-regulated industry or regulated industry in the

---

*Selected for illustration of the non-regulated private economic sector.

Table A-4. Basis for Fixed Charge Rate Annualized Cost Calculations

Technology area	Application area	Financing mode	k_d	k_c	k	N	n	CRF_N	CRF_n	$DPF_{sl,k,n}$	First term FCR[a]
Drinking water	Municipal	100% debt	0.10	-	0.10	20	-	0.118	-	-	0.118
Wastewater	Municipal	100% debt	0.10	-	0.10	20	-	0.118	-	-	0.118
	Non-regulated industry	25% debt 75% equity	0.13	0.14	0.12	20	10	0.136	0.179	0.559	0.162
Particulate control	Regulated industry	50% debt 50% equity	0.14	0.12	0.10	20	10	0.116	0.161	0.621	0.132
	Non-regulated industry	25% debt 75% equity	0.13	0.14	0.14	0.20	10	0.136	0.179	0.559	0.162
Flue gas desulfurization	Regulated industry	50% debt 50% equity	0.14	0.12	0.10	20	10	0.116	0.161	0.621	0.132

[a] First term FCR = $CRF_{k,N} \left[\frac{1 - \tau \times DPF_{sl,k,n} - \alpha}{1 - \tau} \right]$ with $\alpha = 0.10$.

private sector. For non-regulated industry a 25 percent debt, 75 percent equity financing mix was assumed. The regulated industry financing mix was assumed to be 50 percent debt, 50 percent equity. Flue gas desulfurization is primarily a regulated industry (utility) technology; therefore, a 50 percent debt, 50 percent equity financing was assumed.

A.3 Updating Costs

Cost are updated from one base year to a new base year by using cost indices. Cost indices are numbers which reflect relative price levels between different periods in time. A number of different indices are available and are published in a number of references. These include indices for capital cost adjustments as well as adjustments for components that make up operating expenses. The use of cost indices is illustrated by the following equation:

$$\text{new cost value} = \text{old cost value} \times \frac{\text{new cost index value}}{\text{old cost index value}}$$

Cost indices are developed by using a composite of prices for various components that comprise the particular index.

Two capital cost indices used in this report are the:

- *Engineering News Record (ENR) Construction Cost Index.*
- *Chemical Engineering (CE) Plant Cost Index.*

The first is published weekly in *Engineering News Record* magazine. The second is published biweekly in *Chemical Engineering* magazine. Both are publications of McGraw-Hill, Inc. of New York.

In the preparation of this report, costs from existing publications were updated to the reference time of March 1980. Capital costs for water and wastewater treatment were updated using the *ENR Construction Cost Index.* Costs for particulate control technologies and flue gas desulfurization systems were updated with the *CE Plant Cost Index.* March 1980 prices were used for individual cost elements of operating expenses, except for material, which was updated from literature reported values using the *Producer Price Index for Finished Goods* (6).

Table A-5 lists annual averages for the three capital cost indices from 1970 through 1979 as well as end-of-quarter values (last month of the quarter) from March 1977 through March 1980.

A.4 Interest During Construction

The capital costs of a project are paid by borrowing money for the entire project, financing the entire project from internal funds, or financing part of the project by borrowing and part from internal funds. The interest charges associated with construction financing are sometimes capitalized and so can be a significant component of total capital investment.

The interest during construction is the cost of capital required to finance the project during the construction period. Whether the capital is borrowed or internal, there is still a cost for using the capital. For borrowed capital, the cost is clearly the interest charged by a lender. For internal capital, the interest rate is equivalent to the rate of return that the capital could earn if placed in an alternative investment such as a loan to a borrower. Therefore, the cost of capital can be viewed either as interest on a loan or a relinquished rate of return from an alternative project when internal capital is used to finance a construction project. It is anticipated that the chosen project will provide at least that rate of return.

The cost for interest during construction depends on the interest rate, length of the construction period, and schedule of financing during the construction period. A 12 percent per year interest rate was used throughout this report for the construction financing. The construction period was 18 months except for ESP and FF in particulate control and FGD technologies which were 3 months and 30 months, respectively.

Financial schedule becomes important for large projects extending for many months or several years.

Table A-5. Annual Average and End-of-Quarter Capital Cost Indices

	Engineering News-Record Construction Cost Index (1913=100)					Chemical Engineering Plant Cost Index (1969=100)					EPA Sewage Treatment Plant Construction Cost Index (1957-59=100)				
Year	Annual average	Mar.	June	Sept.	Dec.	Annual average	Mar.	June	Sept.	Dec.	Annual average	Mar.	June	Sept.	Dec.
1970	1386					126					144				
1971	1581					132					160				
1972	1753					137					172				
1973	1895					144					183				
1974	2019					165					217				
1975	2209					182					250				
1976	2400					192					262				
1977	2610	2514	2574	2675	2676	204	199	202	209	210	278	271	274	281	288
1978	2811[a]	2698	2822	2851	2872	219	214	218	223	226	305	290	303	311	314
1979	3051	2886	3054	3132	3131	239	237	243	248	239	322	334	338	-	-
1980	-	3150	-	-	-	-	253	-	-	-	-	-	-	-	-

[a] Estimated by averaging end-of-quarter indices.

These projects draw the necessary funds to pay for construction at selected intervals throughout the project rather than all at once. Individual projects have their own specific schedules. For this report, however, for all projects it was assumed that the construciton payment schedule was divided into thirds. One-third of the total funds were required for each third of the toal construction period and were dispersed at the beginning of each period. Another method, sometimes used in utility financing, is to assume that one-fourth of the funds are dispersed during the first and last thirds of the construction interval, and half are dispersed during the second third of the project construction interval. Still another approach includes the assumption that funds are dispersed half way through each third of the construction period rather than at the beginning of each third. This approach will result in lower charges for interest than the assumption that funds are borrowed at the beginning of each load period.

The interest rate and funding schedules discussed above determine the amount of money required for interest charges during a construction project. Table A-6 shows the interest factors for different annual interest rates and construction periods based on the three-thirds schedule discussed above. To obtain the cost of interest during construction, multiply the appropriate factor by total plant cost (see Table A-1).

Table A-6. Factors for Calculating Interest During Construction[a]

Length of construction *period (months)*	Interest rate % per year *10*	*12*	*14*
6	0.03	0.04	0.04
12	0.07	0.08	0.09
18	0.10	0.12	0.14
36	0.21	0.26	0.31

[a]Based on three-thirds loan schedule.

A.5 Location Factors

Construction costs vary geographically due to differences in costs of materials and labor. A sample of this cost variation for wastewater treatment is provided in Table A-7 (7). These construction cost values were derived from calculations using the EPA Sewage Treatment Plant and Sewer Construction Cost Index and should only be used for rough estimates of the geographic influence on capital investment variations for wastewater and drinking water systems. The similarity of drinking water plant construction to that for wastewater treatment plants justifies its use for the former. The labor cost values for plant operating labor were based on a calculation using average earnings from the U.S. Bureau of Census (7).

Similar compilations of factors for particulate matter control and flue gas desulfurization systems were not available at the time this report was written.

Locality factors for power costs by census region rather than major cities are provided in Table A-8 (7).

Table A-7. Cost Locality Factors (7)

	Construction	Labor
Atlanta	0.79	0.77
Baltimore	0.92	0.79
Birmingham	0.79	0.79
Boston	1.04	0.97
Chicago	1.20	1.02
Cincinnati	1.08	0.98
Cleveland	1.13	1.05
Dallas	0.70	0.92
Denver	0.87	1.00
Detroit	1.10	1.32
Kansas City	1.07	0.88
Los Angeles	1.17	1.32
Minneapolis	0.97	1.21
New Orleans	0.94	0.66
New York	1.24	1.14
Philadelphia	1.15	1.05
Pittsburgh	1.02	0.87
St. Louis	1.18	0.83
San Francisco	1.13	1.13
Seattle	1.07	1.21
NATIONAL INDEX VALUES	1.00	1.00

Table A-8. Power Cost Locality Factor (7)

New England	1.31
Mid-Atlantic	1.18
East North Central	1.10
West North Central	0.98
South Atlantic	0.94
East South Central	0.98
West South Central	0.87
Mountain	0.79
Pacific	0.86
U.S. Average	1.00

References - Appendix A

1. Uhl, V.W. A Standard Procedure for Cost Analysis of Pollution Control Operations. Vol. I. User Guide, EPA-600/8-79-018a (PB80-108038*). Vol. II Appendices, EPA-600/8-79-018b (PB80-108046*). U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Research Triangle Park, NC, June 1979.
2. Rudasill, C. Revenue Requirements Calculations for Utility Systems Analysis. Paper presented at the ERDA/METRIC Engineering-Economic Analysis Workshop, the MITRE Corporation/METREK Division, McLean, VA, April 4-5, 1977.
3. Doane, J.W., R.P. O'Toole, R.G. Chamberlain, P.B. Bos, and P.D. Mayevck. The Cost of Energy from Utility-Owned Solar Electric Systems: A Required Revenue Methodology from ERDA/EPRI Evaluations.

---

*Available for purchase from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

JPL Report No. 5040-29 (ERDA/JPL-1012-76/3). Pasadena, CA, June 1976, 82 pp.

4. Bennington, G.E. Ten Steps to Busbar Costs. WP-11488, The MITRE Corporation/METREK Division, McLean, VA, May 1, 1976.
5. Uhl, V.W. Calculations of Annualized Costs Using the Fixed Charge Rate Method. Private Communication. September 1980.
6. U.S. Department of Labor Monthly Energy Review. DOE/IA 0035/05(08).
7. U.S. Environmental Protection Agency, Office of Water and Waste Management, Washington, DC. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009, February 1980.

Appendix B
Glossary

This glossary is presented as an aid to the identification of selected specialized terms which sometimes cause confusion or with which the user might not be familiar.

annual operating expenses: Includes operating and maintenance costs as well as capital related charges except except interest or return on capital. See *expenses*.

annualized cost: The equivalent annual cost equal to the revenue requirement. Includes annual operating expenses plus interest and return on capital.

backwash: In granular media filtration or activated carbon treatment, the reverse flow of clean water through the system to dislodge and remove solids that have accumulated in the bed.

biochemical oxygen demand: A measure of biodegradable organic pollutant content of wastewater expressed as mg/l of oxygen required using a standard test.

capital investment: Investment for long-term use (over a year), which is therefore capitalized.

capital structure: The proportionate portions of capital from sources such as common stock equity, preferred stock equity, and debt (bonds).

capitalize: To consider as an investment; it can either be depreciated (buildings and equipment) or recovered (land or working capital).

cash flow: Annual cash receipts in the form of net profit (after taxes) plus the depreciation charge; also called cash inflow and cash flowback. For comparisons of alternatives with the same revenue, it can be the depreciation charge plus net saving or minus extra net operating charge adjusted for income taxes.

cash flows: The various sources and outlays for funds in an active project.

conceptual estimate: An estimate for a new process or operation, one that has not been built or operated to date.

constant worth dollars: [Current dollars] x [1 + annual inflation rate]$^{-n+1}$, where n is the number of years from the year in question to the reference year. Sometimes these are termed constant dollars, real dollars, or deflated dollars. For an example calculation, see footnote.*

cost index: See *inflation index*.

current dollars: Dollars at any point in time.

depreciation: The allocation in a systematic and rational manner of the cost of fixed capital assests less salvage (if any), over the estimated useful life of the facility.

design flow rate: The flow rate for which equipment is sized. Systems usually operate at less than the design flow rate.

detention time: The residence time of drinking water or wastewater in a process vessel during treatment.

discount rate: The interest rate used either to discount future cash flows to a reference time (zero) or to compound past cash flows to a specified reference time.

discounted cash flow rate of return: See *internal rate of return*.

engineering cost analysis: The application of techniques to the expected capital investments, annual operating expenses, and other cash flows to ascertain the economic feasibility of a project by computing measures of merit.

equivalent annual cost: A generic term to describe equivalent cash flows; it can be calculated either as a uniform end-of-year value, or a uniform continuous flow throughout the year.

equivalent annual value: A version of equivalent annual cost used in evaluating public sector projects.

equivalent uniform cash flow or cost: Corresponds to equivalent annual cash flow (or cost) when it is calculated as a uniform end-of-year value.

escalation: Increase in the cost of a particular item as distinct from general inflation. Escalation might be due to price increases in constant dollars as well as to inflation.

expenses: *Net* expenses are all payments transferred (or paid) to entities outside the operating organization for costs incurred for and related to the plant operation; *total* expenses include depreciation charges in addition to the above.

expensed: The accounting operation in which an outlay is classified as an *expense* and included in an account of expenses, generally classified by type; e.g., operating labor, maintenance materials.

factored estimate: A form of capital cost estimate; usually it is a form of study estimate.

figures of merit: See *measures of merit*.

---

*It is desired to convert the value of an investment of $3 million in 1982 to 1975 constant worth dollars. The inflation rate is estimated at 8% per year. The 1975 value = [$3,000,000] x [1 + 0.08]$^{-7+1}$ = $1,891,000 (1975 constant worth dollars) (then-current dollars).

firing rate: The rate of fuel usage in boilers or other direct-fired process equipment expressed in terms of energy equivalent as Btu/hr.

fixed capital: Corresponds to depreciable investment (buildings and equipment) plus land; excludes working capital.

fixed charge rate: An expression of capital-related fixed charges for a facility as dollars per year.

general expense: An indirectly attributable expense for administration, sales, research, and financing activities.

hydraulic loading: In granular media filtration or activated carbon treatment the flow rate of liquid applied to the granular bed expressed as gpm/ft^2.

I_F: Symbol to denote total plant cost; usually equivalent to the depreciable investment; corresponds to total module cost.

I_w: See *working capital.*

inflation index: Also termed cost index; the relative value of the dollar at a point in time in a particular segment of the economy as compared to its value at an earlier reference time when it is arbitrarily given a value of 100.

interest, continuous: Interest computed by assuming an instantaneous time period for compounding; generally expressed as a nominal interest rate per year. This nominal interest rate works out to be less than the effective interest rate for the year.

interest, discrete: Also termed simple interest; interest on the principal for the period (usually 1 year).

interest rate of return: See *internal rate of return.*

internal rate of return: (IROR) Rate of interest at which outstanding investment is repaid by proceeds of a project to achieve a zero present worth; also, called interest rate of return, discounted cash flow rate of return, and profitability index.

non-regulated industry: An industry in the non-regulated sector of the economy. See *regulated sector*.

measures of merit: Also termed figures of merit, criteria for evaluation, and feasibility criteria; ratios, percentages, and other indices that characterize the economic feasibility of a project; e.g., return on original investment, payout time, internal rate of return, and annualized cost.

minimum acceptable rate of return: This is the lowest return that will be considered attractive for the investment of new capital; it is often taken as the average current return on investment capital; it is not to be confused with the cost of capital and should be somewhat higher. Note that the kind of return (e.g., ROI or IROR) needs to be specified.

module: The major equipment items that carry out either a unit operation (e.g., heat transfer, distillation, solids separation) or a unit process (e.g., biodegradation of liquid wastes).

net annual operating expenses: Operating and maintenance costs as well as capital-related charges except depreciation and interest.

O and M: Direct operating and maintenance costs; represent only a fraction of the total annual operating expenses.

operating flow rate: The flow rate at which a facilty actually operates, as opposed to the design flow rate.

payout time: The time in years to recoup the fixed (depreciable) capital from cash flow; also called payback time or period.

present value: See *present worth*.

present worth: The sum of the discounted (and compounded) values of the cash flows for a given project or operation. The discount rate must be specified.

private sector: Refers to projects financed by private capital and for which the price of the output is set by the market.

reactivation: The treatment of activated carbon to remove adsorbed organic material and restore its adsorption capabilities.

regeneration: Another term for reactivation.

regulated industry: See *regulated sector*.

regulated sector: Refers to projects funded by private capital, but for which the price of the output is regulated by law or a government body. Examples are electric utilities, the telephone company, and public carriers.

reheat: Use of a heat exhanger or reintroduction of some flue gas downstream from a gas scrubber to raise the temperature of the gas to prevent condensation of water vapor in the stack.

retrofit: Equipment or facility added to an existing facility.

retrofit increment: The extra or added cost required for a retrofit facility above that for the basic plant.

total annual operating expenses: Operating and maintenance costs as well as capital-related charges include depreciation but not interest.

total capital investment: The total capital required for a project including various indirect costs such as interest during construction, start-up costs where capitalized, land cost, and working capital.

unit cost: As applied to fixed investment - cost divided by an appropriate output per year; as applied to annual expenses - total expenses divided by output per year; as applied to annualized cost - required revenue divided by annual output. For the latter case, the output may be discounted and escalated in the same fashion as the costs.

working capital: Funds in reserve necessary for the normal conduct of business.

Appendix C
Conversion of English to International System (SI)* Units

	To convert from:	To:	Multiply by:
Length	*ft*	*meter*	*0.305*
Area	*ft*^{2}	meters^2	0.0929
	acres	meters^2	4047
Volume	ft^3	meters^3	0.0283
	gal.	meters^3	0.00379
Mass	lb	kilograms	0.454
	tons	megagrams	0.907
Weight rate of flow	10^3 lb/hr	kg/sec	0.126
	tons/day	kg/sec	0.0105
Vol. rate of flow	gal./min	meters^3/sec	6.309×10^{-5}
	gal./min	millimeters^3/sec	6309
	10^6 gal./day	meters^3/sec	0.0438
Energy	Btu	kilojoule (= Newton x meter)	1.055
	kW-hr	megajoules	3.60
Power	hp	Joules/sec	746
	kW	Joules/sec	1000
	10^6 Btu/hr	kilojoules/sec	293
Specific energy	Btu/lb	kilojoules/kg	2.324
Pressure	lb/in.^{2}	kilopascal (= kilonewton/m^2)	6.895
Water for energy	gal./10^6 Btu	m^3/megajoule	3.592×10^{-6}
Heat rate	Btu/kW-hr	Joules/kW-sec	0.293
Temperature	°F	K	0.556 (°F + 459.7)
Heat transfer Coefficient	Btu/hr ft^2 °F	Joules/secm^{2}K	5.674

**Standard for Metric Practice*, Amercian Society for Testing and Materials, E3 80-76, 1976.

Appendix D
Miscellaneous Conversion Factors

Multiply	By	To Obtain
acres	43,560	ft^2
atmospheres	29.92	in. of mercury
atmospheres	33.90	ft of water
atmospheres	14.70	psi
Btu	1.055	kJ
Btu	777.5	ft-lb
Btu	3.927×10^{-4}	hp-hr
Btu	2.928×10^{-4}	kW-hr
Btu/lb	2.326	kJ/kg
cu ft	28.32	liter
cu ft	0.03704	cu yd
cu ft	7.481	gal.
cu ft/second	0.6463	mgd (million gal./day)
cu ft/second	448.8	gpm
cu yd	0.765	m^3
°F	0.555 (°F - 32)	°C
ft	0.3048	m
gal.	3.785	liter
gal., water	8.345	lb, water
gpd/sq ft	0.04074	$m^3/m^2 \cdot day$
gpm	0.06308	liter/s
gpm/sq ft	0.06790	liter/$m^2 \cdot s$
hp	0.7457	kW
hp	42.44	Btu/min
hp	33.00	ft-lb/min
hp-hr	2.685	MJ
in.	25.4	mm
lb (mass)	0.4536	kg
million gal.	3,785	m^3
mgd (million gal./day)	3,785	m^3/d
ppm (by weight)	1.000	mg/liter
psi	6.985	kN/m^2
sq ft	0.0929	m^2
tons (short)	907.2	kg

Conversions between MWe, firing rate, and gas flow depend on fuel and excess air used. For this report the factors used for particulate matter control and wet FGD systems are:
3200 acfm = 1 MWe
412 acfm = 10^6 Btu/hr
For spray drying FGD the factors are:
3400 acfm = 1 MWe
360 acfm = 10^6 Btu/hr

☆ U.S. GOVERNMENT PRINTING OFFICE: 1984 - 559-111/10719